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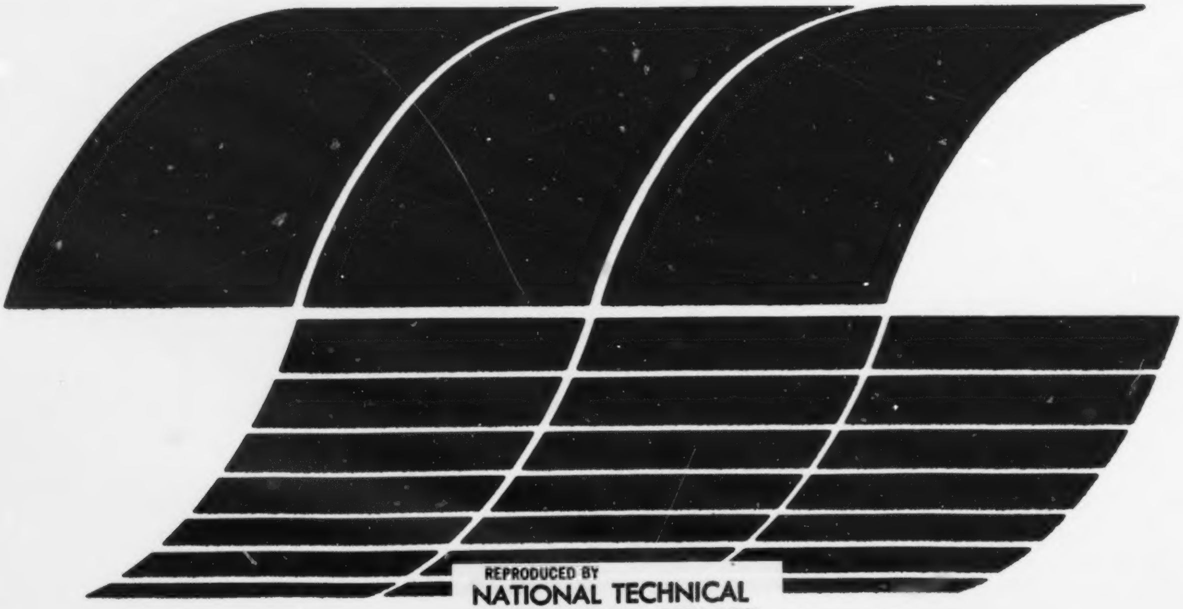

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Research Triangle Institute
P. O. Box 12194
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Office of Energy, Minerals, and Industry
Research Triangle Park, NC 27711

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SESSION 1

ENERGY AND THE ENVIRONMENT

RICHARD D. STERN, CHAIRMAN

Overview of Control Technology - The Bridge Between Energy Utilization and Environmental Goals

BY

Frank T. Princiotta
Director, Energy Processes Division
Office of Energy, Minerals, and Industry
U.S. Environmental Protection Agency

AND

Clinton W. Hall
Director, Energy Coordination Staff
Office of Research and Development
U.S. Environmental Protection Agency

Energy Policy - Where Are We Going?

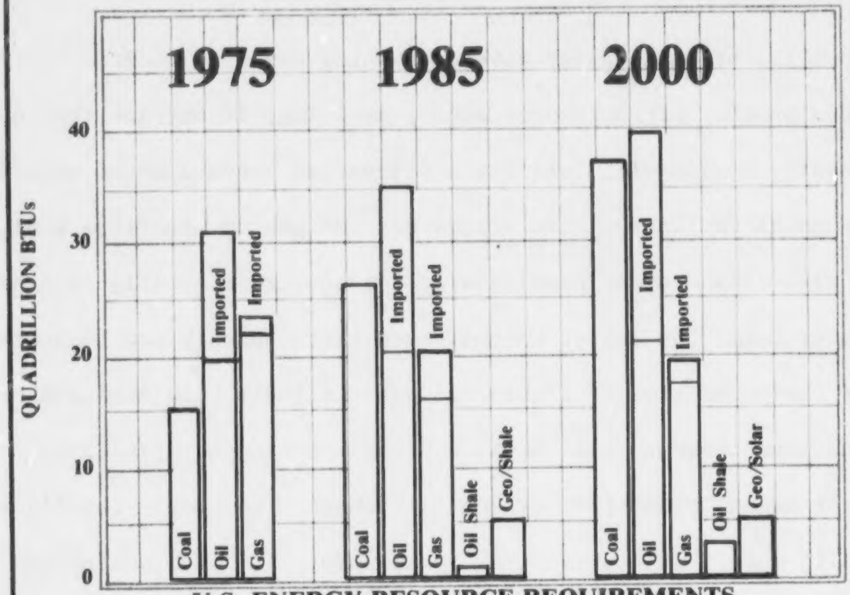
The profile of U.S. energy development and use will undergo major changes in the years ahead. Although only slowly evolving, it appears that our national energy policy will call for a widespread conversion of utility and industrial power facilities from scarce oil and gas to plentiful coal, decreased fuel consumption, particularly for the transportation sector, and, in the longer term, the use of technologies that are only now beginning to emerge for the production of liquid and gaseous fuels from coal and oil shale.

Projections indicate that total U.S. coal mining activities will increase from today's annual production of 700 million tons to nearly 1 billion tons by 1985 and will more than double by the year 2000. In 2000, conversion of existing utility and industrial facilities from oil and gas to coal coupled with construction of new conventional and advanced coal utilization facilities will consume approximately 1.4 billion tons of coal annually (Figures 1,2, and 3). Although conventional combustion of coal will predominate, by the year 2000 emerging coal-based technologies are projected to consume 300 million tons of coal per year. Furthermore, the National Highway Traffic Safety Administration predicts that diesel powered cars, which offer 20 to 30% gains in fuel efficiency, will account for 25% of all new car sales in 1985 (Figure 4).

Environmental Problems

These shifts in energy development and use pose potential significant threats to human health in the next two decades. Massive increases in coal and oil shale mining, off-shore oil and gas production, and uranium extraction are all projected by the year 2000. Intensified mining activity will create

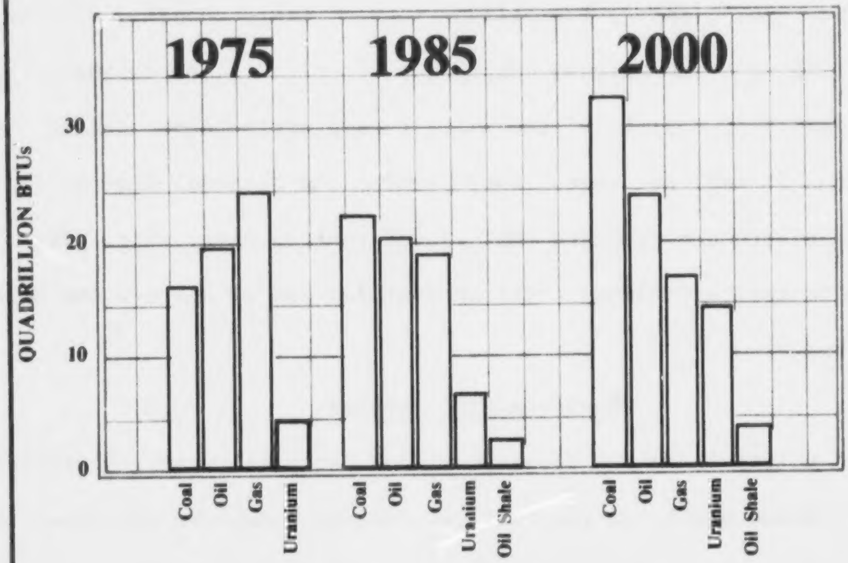
Figure 1



U.S. ENERGY RESOURCE REQUIREMENTS

Source: Technology Assessment Modeling Project, 1978

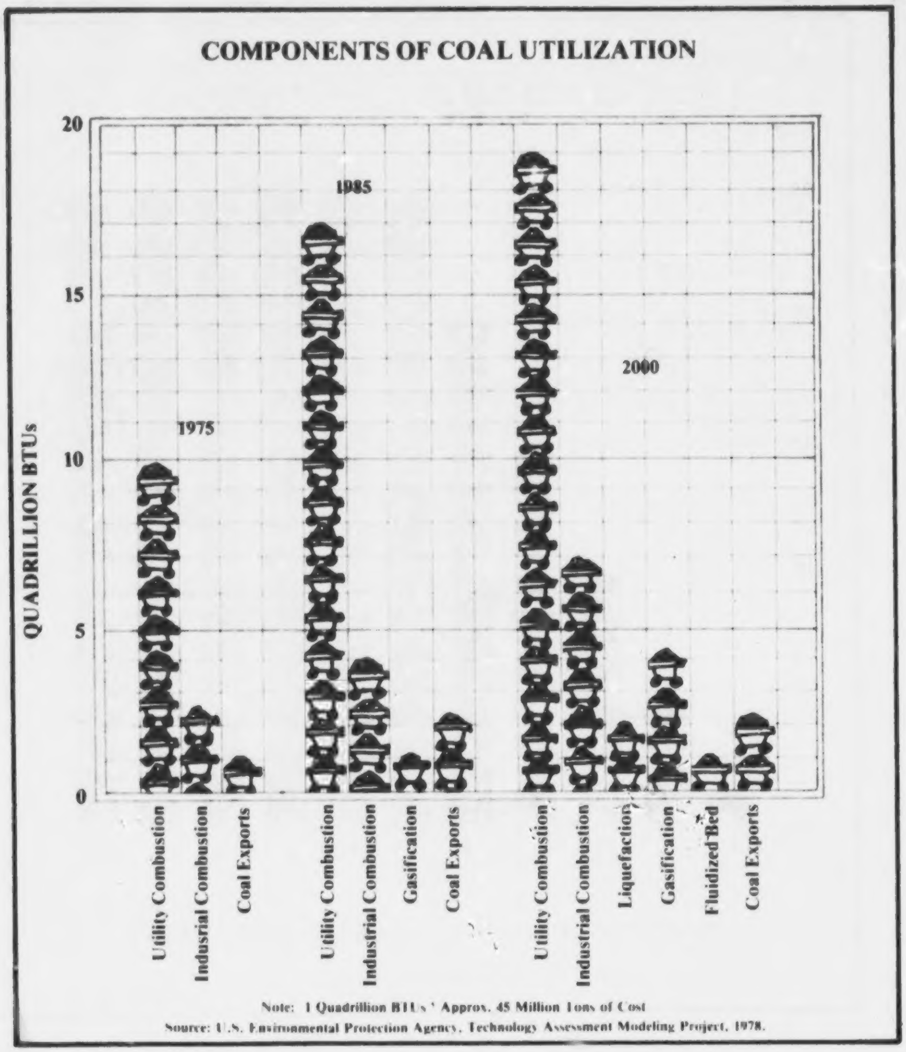
Figure 2



SOURCES OF DOMESTIC ENERGY SUPPLY

Source: Technology Assessment Modeling Project, 1978

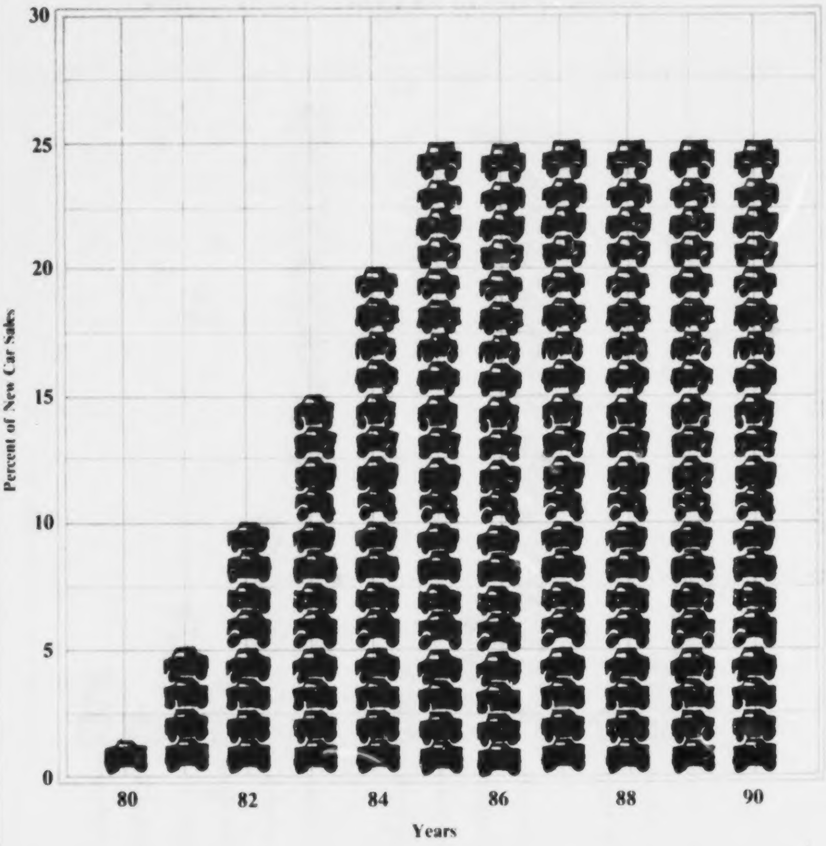
Note: 1 Quadrillion BTUs = Approx. 45 Million Tons of Coal



Note: 1 Quadrillion BTUs = Approx. 45 Million Tons of Coal

Source: U.S. Environmental Protection Agency, Technology Assessment Modeling Project, 1978.

Figure 3



MAXIMUM PROJECTED DIESEL AUTO SALES

Source: National Highway Traffic
Safety Administration

Figure 4

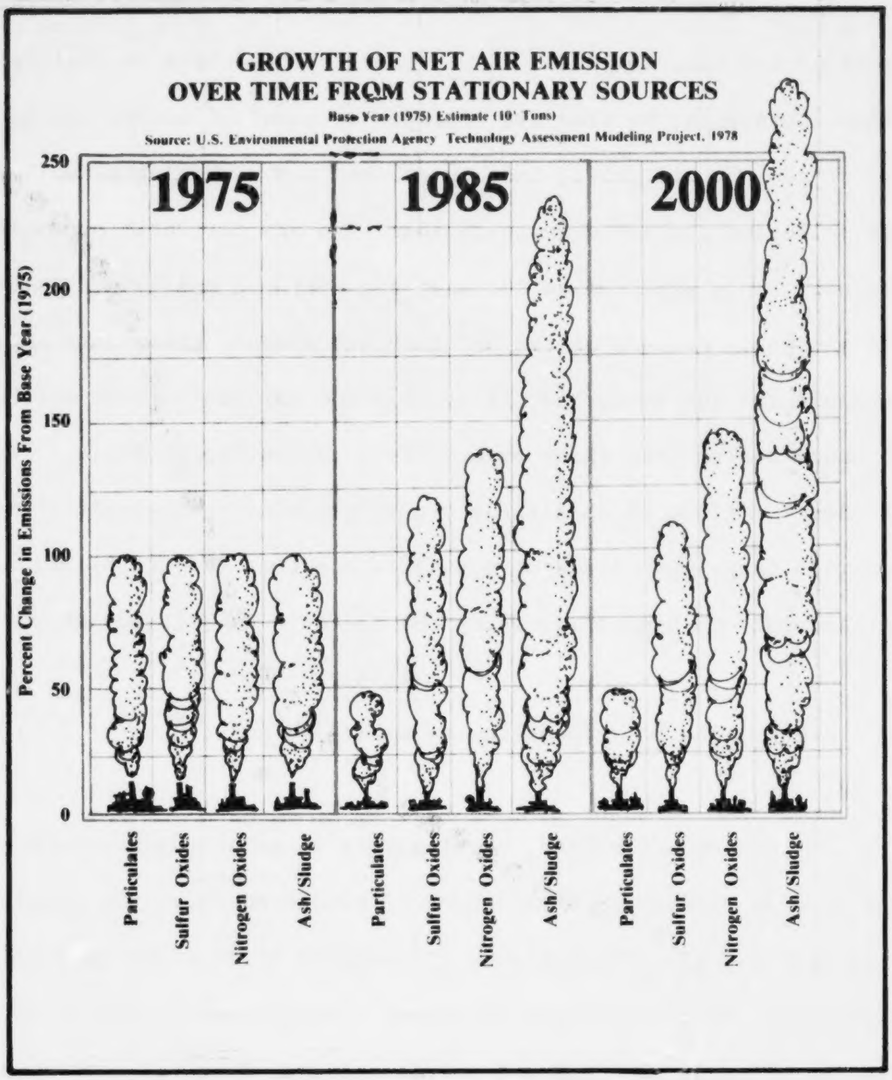
erosion problems and generate runoff which can contaminate surface waters. Aquifers may also be polluted as a result of leachate or drainage from the mines themselves, or from the improper disposal of mining wastes. Increased use of coal by utilities, industries, and new technologies will produce more air pollution and solid waste residue than are currently produced (Figure 5). The pollutants expected to increase are nitrogen and sulfur oxides, and ashes and sludges. Because of the way they are formed, pollutants emitted from new technologies can be varied and complex and may prove to be even more harmful to human health than those emitted from current technologies. And the use of diesel engines as an alternative to gasoline spark-ignition engines will generate large quantities of particulate matter (Figure 6) which may be carcinogenic to humans and which EPA research has already shown to be mutagenic.

Research Needs

A multitude of information is needed to avert massive future environmental impacts.

For mining activities, particularly those associated with coal, oil shale, and uranium production, the impact of runoff on receiving streams and of mine drainage of toxic pollutants on groundwater needs to be quantified and the appropriate control methods developed. Techniques to combat water and wind erosion of reclaimed land are sorely needed. Improved methods are also required to mitigate radiation problems resulting from mining and milling uranium ores.

Expanded use of existing coal burning technologies demands that technologies for sulfur oxide and particulate control be improved (Table 1). Since they are in such an early stage of development, control technologies for nitrogen oxides must undergo vast improvement in the years ahead. Additionally, research must

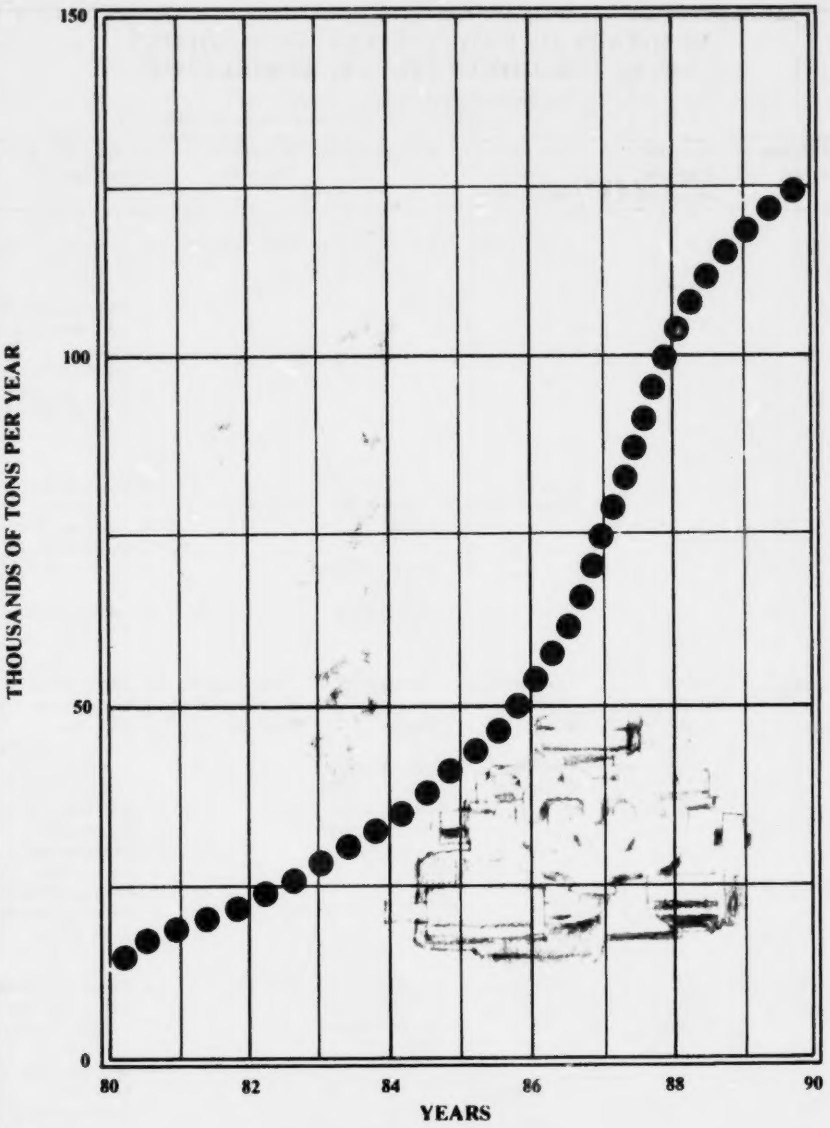


GROWTH OF NET AIR EMISSION OVER TIME FROM STATIONARY SOURCES

Base Year (1975) Estimate (10^6 Tons)

Source: U.S. Environmental Protection Agency Technology Assessment Modeling Project, 1978

Figure 5



DIESEL PARTICULATE EMISSIONS

Source: Office of Research and Development U.S. Environmental Protection Agency, 1978

Figure 6

Table 1

SUMMARY OF R&D CONTROL TECHNOLOGY NEEDS FOR CONVENTIONAL COMBUSTION

Source: Princiotta, F.T., 1977, Utility and Industrial Power, Energy/Environment II.
U.S. Environmental Protection Agency EPA-600/9-77-012 (Updated for Research Outlook, Aug. 1978)

Description of Pollutant	Standard Presently Established	Type of Control Technology	Present Status	Secondary Residuals	Needed Control Technology R&D
Sulfur Dioxide (SO_x)	Yes NSPS & AAQS*	Coal Cleaning	1st Generation Demo Planned	High S-Refuse	— Eliminate Secondary Pollution — Demonstrate Practicability in Conjunction with FGD — Develop Chemical Processes Capable of High Efficiencies
		Flue Gas Desulf. (FGD)	1st Generation in Full Scale Demo 2nd Generation in Bench and/or Pilot Scale	Sludge, Purge Streams	— Improve Removal Efficiency — Eliminate Secondary Pollution — Improve Reliability — Improve Energy Efficiency
Nitrogen Oxides (NO_x)	Yes NSPS & AAQS*	Combustion Modification (CM)	Commercial for Some New Units	Purge Streams for Certain Processes	— Lower Costs — Demonstrate Low NO_x Burner Capable of 80% Control — Broaden Applicability of Combustion Modification Technology — Evaluate Flue Gas Cleaning Process at Pilot Scale
		Flue Gas Treatment	Pilot Scale and Demo in Japan on Oil; Pilot Scale on Coal in U.S.		
Particulate Matter	Yes NSPS & AAQS*	Electrostatic Precipitators	Commercial- 1st Generation Demo	Fly Ash	— Improve Conventional Fine Particulate Control Technology — Broaden Applicability — Develop Novel Devices with Improved Capability and Cost Effectiveness
		Bag Houses	1st Gen. Commercial		
		Wet Scrubbers	2nd Gen. Full Scale Demo		
		Novel Devices	Bench or Pilot Scale		
Potentially Hazardous Materials	No	Undefined	Undefined	Undefined	— Assess the Magnitude of Problems Associated with Unregulated Pollutants Via Chemical and Biological Characterization

* Ambient Air Quality Standard (Health-Related)

strive to address major information gaps regarding the magnitude of the health and environmental problems associated with trace elements, radioactive material, and polycyclic organic emissions produced during conventional combustion of coal.

Emerging energy technologies must also undergo careful scrutiny for environmental impacts (Table 2). Particularly, the level of sulfur and nitrogen oxides, particulates, heavy metals, and toxic and carcinogenic organic compounds, produced by the coal-based technologies of gasification, liquefaction, and fluidized bed combustion need to be assessed. Geothermal energy production methods should be examined from the perspective of hydrogen sulfide gases released, waste heat and steam plumes produced, and land-use implications. Solar energy systems should also be evaluated in terms of land and water use, sludge and residual production, and the possible leakage of toxic working fluids.

Finally, research needs to determine the cancer-causing potential of diesel soot and, if positive results are found, to establish the link between ambient concentrations and the incidence of cancers in humans. In parallel, control technologies will be evaluated and developed which offer potential for major reductions in particulate emissions.

The Bridge - Control Technology

So we see an evolving national energy policy which through its emphasis on increased coal combustion, conservation and emerging energy technologies can yield massive environmental damage. At the same time national concern for environmental quality remains high and stringent legislation is on the books which calls for protection of our air, water and land resources.

Table 2

SUMMARY OF EMERGING ENERGY TECHNOLOGIES

Emerging Energy Technologies	Status	Major Environmental Concerns
FOSSIL FUEL		
Coal Gasification	Lurgi low BTU gasification is commercial in Europe for certain non-coking coals. Various low BTU processes are and will be demonstrated in the United States under Department of Energy funding. Low BTU gasification for on-site heating will be available in mid-1980's. Methanated Lurgi and second generation high BTU process will be available in 1990's.	Sulfur and Nitrogen compounds Particulate emissions Water contamination Heavy metals and organic compounds Acidic gases Subsidence (*in-situ* gasification) Aquifer disruption (*in-situ* gasification)
Coal Liquefaction	Major processes near commercialization are Solvent Refined Coal (SRC) and H-coal processes. Currently at pilot scale; demonstration of two processes planned by the Department of Energy.	Sulfur and nitrogen compounds Particulate emissions Water contamination Heavy metals and organic compounds Acidic gases
Coal Fluidized Bed	30MW pilot FBC (atmospheric) being studied at Rivesville, West Virginia by the Department of Energy. Availability expected in 1990. Small pressurized EPA unit has been operated for several years (0.63MW); DOE plans a 14MW pilot unit in the early 1980's. Availability expected in 1990's.	Sulfur and nitrogen oxides and particulates Toxic heavy metals and organic compounds Thermal pollution Spent sorbent disposal problems
Oil Shale	Both above ground and underground (in situ) retorting processes under development. Major on-going efforts include the Navy's Program, using the Paraho above ground process, nearing completion of its goal to produce 100,000 barrels of shale oil and Occidental *in situ* process which has produced in excess of 50,000 barrels to date.	Sulfur and nitrogen compound emissions Particulate emissions Water contamination and availability Overburden and spent shale Toxic and combustible gases Subsidence and aquifer disruption *(in situ)*
OTHER		
Geothermal	Three principal types: Convective hydrothermal; geopressurized hydrothermal and hot dry rock. Present generating capacity of convective hydrothermal is 500MV. Hot dry rock is the largest resource but because of the difficulty in fracturing the rock, it has generated no commercial interest. Increased geothermal application is expected in the 1990's.	Hydrogen sulfide release Waste heat and steam plumes Seismic effects Subsidence Minerals precipitation Noise and blowout Land use
Solar	Three major areas: Heating and cooling of buildings, production of electricity (photovoltaic) and production of clean fuels from biomass. Clean fuels at commercial scale from biomass (gasohol) are currently being produced; and space heating currently state of the art.	Toxic working fluid leakage (photovoltaic) Sludge and residuals from silicon distillation (photovoltaic)

In order to resolve this potential conflict, it is essential that economical and reliable environmental control technology be developed and ultimately applied on a widescale basis. Control technology, then, allows the nation to meet two of its major goals - adequacy of reliable energy supplies and environmental protection.

REMARKS BY LEON RING, GENERAL MANAGER
TENNESSEE VALLEY AUTHORITY
BEFORE SYMPOSIUM ON FLUE GAS DESULFURIZATION
LAS VEGAS, NEVADA, MARCH 5, 1979

I appreciate this opportunity to meet with you. I know this group recognizes the need for technical exchange in the area of environmental control. And I believe that technical exchange will occur. On your program, I see buyers, sellers, builders, operators, researchers, and regulators: I see contributors from the United States and abroad. And I'm sure public interest environmental organizations are well represented also.

This type of forum is especially appropriate for indulging in frank discussion, bringing problems to the forefront, and finding ways to resolve them. One area I'm sure you will discuss this week is flue gas desulfurization system reliability. As more and more systems are put into use, improvements will be essential for the success of this technology. Many other things will be discussed here. As a matter of fact, it's entirely possible that future regulations may be shaped from discussions held at this meeting. It's possible, too, that company plans may be amended due to exchanges here in Las Vegas. Whatever area the impact is in, I'm sure it will be constructive. Some of us who indulge at the gaming table may lose a few coins, but I think we'll leave here winners with new information and new ideas toward solving environmental problems.

I guess I ought to warn you that I'm the first of five TVA speakers. The others have told me not to cover their material. We would hope to be able to provide some insight into every area of TVA's FGD program. My remarks will be more general and touch on TVA's policy in environmental control. Others will tell you of our findings, our experiences, and our detailed plans to clean up the air in the Tennessee Valley. In fact, just before the meeting today Jerry Hollinden told me that he was prepared to expound on any detail of our 50-page proposed consent decree. So just shoot him the page, paragraph, and line.

I also hope that our participation at this symposium shows our support for this type of forum and our deep interest in a subject that relates so significantly to our nation's environmental health. As you may know, TVA has launched the country's largest program to clean the air and we are fully committed to making it work. TVA is more than a power system. We have broad responsibilities for the economic and social development of a vast region. Clean air is consistent with our purposes and an integral part of improving the quality of life in the Valley.

TVA was the first experiment in unified development of the total natural resources of a river valley, pioneering an idea that has since spread far and wide. TVA has built dams to regulate the Valley's floodwaters. The system of dams also created a waterway for barge traffic and generates electricity. TVA develops and demonstrates improved fertilizers which have helped to replace erosive row crops on hillsides with pastures and cattle. Our role in fertilizer development is a little known fact, yet half of the fertilizer research in the world is done by TVA. TVA also works with the states and other organizations in developing agriculture, forestry, recreation, and other resources.

Protection and enhancement of environmental quality is and has always been an important part of TVA's concept of integrated development of the resources of the region. This is no small job in the Tennessee Valley. Protecting the environment in our 80,000-square-mile service area has become a complex and demanding task when coupled with the expansion of power generating capacity.

The TVA power system, right now, consists of about 1,200 MW in pumped hydro storage, 2,500 MW in combustion turbines, 3,300 MW in hydro capacity, 3,500 MW in nuclear capacity, and 18,000 MW in coal-fired capacity. Under long-term contracts, we have access to an additional 1,300 MW in hydro capacity. With some quick addition, you'll find that's a total of close to 30,000 MW of installed capacity. Our present construction schedule calls for an additional 18,000 MW to be in service by 1986. Most of that will be nuclear--the exception is one more unit of pumped hydro.

Forecasting growth in power loads used to be done in a simple way--often with no more effort than drawing lines on log-log paper. But the lead time now required for additional capacity and the number of factors affecting future loads have made more sophisticated forecasting mandatory. So TVA has gone to an extremely complex forecasting methodology.

A number of interacting factors are avaluated for their impact on future loads. Among them are such factors as economic growth, population growth, and substitution of electricity for scarcer fuels, which tend to add to power demands. Others such as conservation, load management, the use of solar energy and congeneration, increased prices, and appliance efficiency will make the forecasted load go down. All these things complicate making estimates on how much capacity we will need and by when.

By quantifying these factors in ranges, our power supply planners come up with a wide range of forecasts. Load forecasts for the next 10 years project that between 200 billion and 230 billion kilowatt-hours will be sold in 1988 And it seems clear that we will need additional capacity in the 1990s.

But TVA owes more than adequate power to the 6.7 million Tennessee Valley residents. We owe them a clean and safe environment, and we intend to provide one. In 1950, even before the first of our steam plants went into operation, TVA began an extensive program of air quality studies and monitoring. That program has provided some of the best basic data available for predicting the effects of power plant operation on air quality under various weather conditions. This information has been valuable to TVA and other systems in locating, designing, and operating power plants.

In the 1960s and early 1970s TVA went about the business of protecting the environment by using a variety of methods. We built tall stacks as much as 1,000 feet high to disperse pollutants; we installed electrostatic precipitators to control fly ash; emissions were limited by operational control procedures (we called this method SDEL). Selective use of clean fuels and selective installation of SO_2 removal systems were practiced or planned.

A master plan that would enable TVA to meet ambient sulfur dioxide and particulate emission standards at all of its plants was proposed in 1973: This plan included the use of tall stacks at three plants; intermittent controls at nine plants; and an experimental SO_2 scrubber at the Widows Creek plant. At that time, TVA felt that ambient sulfur dioxide standards fully protected the public welfare from any known or anticipated effect of sulfur dioxide. And there was doubt that continuous SO_2 emission standards were necessary to meet the requirements of the 1970 Clean Air Act for existing installations. Using these methods, TVA could have met ambient standards at about one-tenth the cost of either utilizing scrubbers or low-sulfur coal.

This TVA approach differed significantly from what EPA proposed. This difference revolved around interpretation of what was required under the Clean Air Act and was finally settled by the Supreme Court in 1976 when they refused to hear our case. This decision finally settled the question and necessitated the use of continuous emission control. And with that decision TVA moved into a new era of sulfur dioxide emission control.

Since the Supreme Court ruling, TVA has faced litigation in several different courts. A coalition of citizens groups filed suit against TVA in 1977. The States of Kentucky and Alabama and EPA also entered the suit to force TVA to meet SO_2 emission requirements. These suits have been consolidated and are awaiting trial. But TVA expects them to be settled with the approval of TVA's strategy for compliance. The TVA Board and I have already approved this plan. The settlement is now subject to approval by Federal courts in Nashville and Birmingham.

The agreement calls for the use of low- and medium-sulfur coal, coal washing, and more scrubbers, and will bring all TVA plants into full compliance with requirements of the Clean Air Act.

Carrying out the settlement will not be without problems. The massive amounts of construction work that must be scheduled between now and 1982 are staggering. Scheduling unit outages for scrubber tie-in while maintaining power loads will be another challenge for TVA's power system operators. And when we get all those FGD systems on line, operating problems will come in multiples rather than singly as they do now from our Widows Creek scrubber.

We realize that carrying out the settlement will also be expensive. Our real capital investment will be in the neighborhood of a billion dollars. (That's a ritzy neighborhood.) Operating expenses (in 1982 dollars) are estimated at over $400 million per year. Yet, while there is an obvious impact on residential customer rates, electric bills will climb gradually, reaching a maximum increase of about 9 percent in 1983 and then will decrease in the succeeding years.

The benefit of the settlement is obvious--a cleaner environment for the Tennessee Valley. More than 970,000 tons of SO_2 and 85,000 tons of particulates a year will be removed from the skies of the Tennessee Valley due to the pollution controls under this agreement.

The benefits of this program will not be limited to the Tennessee Valley though. As we have in the past, TVA plans to share the knowledge gained from our experience with you and the organizations you represent. We look for continued and increasing participation with EPA, EPRI, DOE, and EEI in the area of technical exchange. I'll mention a few examples of this type of work but I'll look to my colleagues in TVA to give you details in their papers later in the symposium.

Despite our sometimes conflicting views, TVA and EPA have made some outstanding joint contributions in FGD research. I can cite examples like our Shawnee test facility, where we developed and demonstrated FGD design and operating practices and found possible sludge oxidation techniques for producing a better disposal product. The list is long and the accomplishments substantial.

Our work with EPRI is no less outstanding. The cocurrent scrubber research that we have undertaken for many years together has provided us with valuable data on a novel SO_2 scrubber device. And we are planning additional projects in the next few months. With the recent appointment of TVA's Chairman to the EPRI Board, our longstanding cooperative relationship should be enhanced.

It has been more recently that TVA has stepped up its participation with DOE. Under a Memorandum of Understanding signed in April 1978, TVA and DOE have begun many joint projects. Several of these deal with the environment and its protection. In addition, TVA and EEI have long been involved in the exchange of technical information.

Let me turn now to another angle of environmental protection--not by cleanup approaches but by prevention. TVA is involved in demonstrations of those energy technologies with minimal or reduced environmental impact. I think you will be interested in knowing how involved we are in such areas as fluidized bed combustion, solar energy, fuel cells, and waste heat utilization.

Atmospheric fluidized bed combustion holds potential for dramatic improvements in SO_2 emission levels over conventional coal-fired boilers. It may allow a greater use of the predominately high-sulfur coal found in the eastern United States. For thise reasons, TVA has taken a lead role in developing AFBC and is planning to build a 200-MW unit in the mid-1980s. Preliminary design is essentially complete for this 200-MW boiler. In support of this program, we will build a 10-MW pilot plant.

To address the ever-growing problem of solid waste disposal and its environmental impact, TVA is pursuing advancements in the technologies that recover minerals and energy from waste. TVA and the city of Gallatin, Tennessee, have begun a five-year development of a $7.9 million solid waste cogeneration facility. TVA is providing both financial and technical assistance. This type of program is also being considered for other areas of the Tennessee Valley region.

TVA views its solar program as a real winner in the environmental area. A demonstration project involving the planned installation of 1,000 solar water heaters is now being implemented in Memphis. TVA has arranged financing for these units, providing credit for restricting the electric hookup system to off-peak use.

Another area where TVA has made significant progress is in our home insulation program. TVA, through its distributors, surveys homes, makes recommendations on conservation measures needed, and provides low-interest financing for the measures. Over 98,000 homeowners have already been surveyed. We are now moving to expand this type of program to commercial and industrial customers. We believe if we can reduce the amount of electricity we have to produce, we have reduced the associated pollution.

The TVA staff has recently assessed the outlook for fuel cells placed at dispersed sites throughout the Valley. We are looking at a system that would use gasified coal, distributed by pipeline from mine-mouth operations. The outlook for this approach is promising. Coal use efficiency should be high and environmental effects should be minimal.

Ways to use the low-grade heat rejected in normal power plant operation are being explored by TVA. TVA has operated a simulated waste heat greenhouse in Muscle Shoals, and has grown catfish in waste heat water at one of its plants. A greenhouse at the Browns Ferry Nuclear Plant is heated with waste heat water from that plant's condensers. The results to date have been positive to the point that we are planning a waste heat industrial park at the Watts Bar Nuclear Plant. This park would bring agricultural, aquacultural, and industrial users of waste heat together--a first for the United States. Operation of this innovative park should begin in 1982.

Taking this one step further, TVA is also assessing the feasibility of building and operating a central cogenerating power plant to provide power for the TVA grid and to supply intermediate- and high-quality process steam to industries. We have received inquiries from industrial customers about this arrangement and we are encouraged by our discussions with them on cogeneration's potential.

We at TVA do not view environmental protection and energy technology development as opposing forces. In our view they are two sides of the same coin. We must have energy technology development to protect our environment adequately. TVA will pursue a course that will enable the people we serve to have a clean environment and energy assurance.

Health Effects of SO_2 and Sulfates

by

Bertram W. Carnow, M.D.
Professor
Occupational and Environmental Medicine
School of Public Health
University of Illinois at the Medical Center
P.O. Box 6998, Chicago, Illinois 60680

and

Edward Bouchard
Research Assistant
Environmental Health Resource Center
School of Public Health
University of Illinois at the Medical Center
P.O. Box 6998, Chicago, Illinois 60680

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INTRODUCTION

In preparing my discussion I had to ask myself the question, "Why was I being asked to discuss the health effects of SO_2?" Along with particulates, it was the earliest of the common pollutants studied to determine its health impact on the community. Our first study in the early 60's along with a number of others used in setting the SO_2 standard revealed that increased respiratory and cardiac deaths occurred at elevated levels of SO_2. Other studies also revealed increased episodes of acute illness in those with chronic disease as well as acute health effects on other segments of the population. In 1973 I participated in a three (3) day conference held under the aegis of the National Academy of Sciences. We concluded that more studies were needed regarding health effects, but that there was no basis for lowering the standards for SO_2 or indeed for any of the others. A similar conclusion was arrived at by a National Academy of Science Task force on multiple pollutants. The task force on SO_2 which I chaired, again arrived at this general conclusion as did a study carried out by Dr. David Rall of NIEHS for O.M.B.

The problem,I believe, is that many studies have been carried out on animals and humans with variable results regarding the level of effect. Since we will need to rely on coal as a major energy source and since SO_2 removal is costly, the variability in results is confusing and thus the related question, "Is SO_2 really bad for health?" is again raised. As you will hear from my discussion, it is my firm conclusion based on clinical experience and epidemiologic studies carried out by our group and others that it may indeed adversely affect health and in some marginal groups in the population be life threatening. A number of studies we just completed and which I will discuss briefly, tend to confirm this conclusion.

The confusion regarding results I believe extends from the following:

1. Experiments have been carried out on a multitude of animal species which have variable sensitivity to SO_2.

2. The doses used and the duration of exposure have also varied greatly.

3. Other conditions in the laboratory have also been highly variable including variations in the age of animals, temperature, and humidity, all of which may be important factors in assessing effect.

4. The experiments on animals and humans generally were carried out using SO_2 as the gas in the laboratory, whereas the major effect, certainly the most severe and chronic effects of SO_2 are due to sulfate and sulfuric acid, secondary pollutants formed from SO_2.

5. Human experimentation in addition to being carried out under circumstances where no other pollutant is present, are usually carried out on healthy, young adults. Ethical considerations prevent us from using those individuals at highest risk such as those with chronic heart and lung disease or asthma.

In view of this, my presentation should more appropriately be titled "Health Impact of SO_2" because in the real world it is found along with many other pollutants including particulates, ozone and others with which it may act synergistically. Further, it becomes particulate and aerosol after interacting with other materials in the environment and therefore, what we may be seeing are the effects of the off-spring and not of the parent. Possibly most important of all is the fact that all of these pollutants including SO_2 and those with which it interacts are impacting on a heterogenous population which includes the very young, very old with chronic diseases of the heart and lung and other organs, millions with genetic predisposition due to allergies, asthma, alpha 1 anti-trypsin deficiency and others. In addition, 70 million Americans are at high risk because of cigarette smoking. These are all more susceptible.

I. HEALTH EFFECTS OF SO_2 AND SULFATES

SO_2 is classified as an irritant gas, in that it produces irritation and inflammation of the tissue that it contacts directly. SO_2 increases flow resistance by constriction of the airways, decreases the elastic recoil of the lung slightly, and, at higher concentrations, can decrease breathing frequency. SO_2 may also be absorbed by the blood and has some effect through a central nervous system mechansim. Additionally, chronic exposure may result in decreased mucociliary flow, a major defense system which I will discuss. Sulfuric acid is also classified as a primary irritant in that its irritant and corrosive action far exceeds any systemic toxic action.

II. GENERAL PHYSIOLOGIC AND ANATOMIC CONCERN

The lung has multiple defenses against environmental assaults. The nose and upper respiratory tract warm, filter and humidify the air. Chemoreceptors in the nose and airways can detect irritant gases and produce sneezing, coughing, bronchial constriction and narrowing, and other reactions to prevent noxious material from reaching deep lung areas. Lining the airways of the lung are cells with tiny hairs attached (cilia) which sweep invading particles upward in waves at the rate of 1500 times per minute. Furthermore, in response to irritation, the body produces mucus. The mucus blanket lining the air passages trap foreign particles. The cilia then move the mucus blanket up and out, in an escalator-like motion, thus removing the particles from the lungs. This clearance mechanism is a primary defense of the lung. In a normal healthy individual, most of the larger particles are trapped before they reach the deeper, more vulnerable parts of the lung. Particules from two to five microns and below in size, however, can penetrate to the alveoli. (air sacs) Within the alveoli, particles can be removed by scavenger cells (macrophages) which contain digestive enzymes such as trypsin which attempt to consume the particle. Trypsin is itself highly toxic however, and its release may be related to alveolar destruction or the development of emphysema (Bates, 1972). People who do not have the ability to neutralize trypsin are extremely vulnerable to all forms of air pollution (Bates, 1972).

The defenses of the lung can fail to operate for various reasons such as aging, illnesses, genetic effect, or simply being overwhelmed by a toxic substance. When it is overtaxed for any reason the respiratory system is even more vulnerable to injury from either acute or chronic exposure to environmental pollutants like SO_2 (Cassarett, 1975). For example, in aging, the capacities of the respiratory system gradually decrease (Morris, 1971). Smoking and/or air pollution can speed up this aging process considerably (Bates, 1972). Thus certain older populations, such as adult males over 55 with chronic bronchitis, have been identified at high risk during air pollution episodes (Carnow etal 1968 Carnow and Feiveson, 1969). Infants and children whose respiratory defenses are not fully developed are similarly vulnerable (Goldstein, 1975).

With respiratory tract impairment it takes more work to breathe. As a result, there is an increase in lung pressure, which places strain on the right side of the heart. This may enlarge the right side of the heart, a condition known as cor pulmonale. Additionally, aggrevation of cardiovascular diseases, particulary coronary artery disease, have been associated with high levels of SO_2 as a result of a reduction in available oxygen (Carnow, 1973).

III. ANIMAL EXPERIMENTAL STUDIES

Many studies have been carried out on multiple species. These are extensively reviewed in the literature and summarized in table 1 in the text. Many other summaries are also available. I would only like to summarize Dr. Amdur's conclusions after many years of studying the toxicity of the aerosols formed by the oxidation of SO_2.

Table I Summary of Selected Experiments with Animals Exposed to SO_2 Alone or in Combination with other Pollutants (Sulfates, acid mists and virus)

Type of Animal	SO_2 Concentration	Exposure Time	Type of Test	Results	Reference
Mice	10 ppm	24,48 & 72 hrs.	In Vivo	Severe injury manifested more in the nasal cavity than in the trachea or lungs.	Giddens and Fairchild, 1972
Mongrel dogs	7-230 ppm	15-20 min.	In Vivo	Changes in flow resistance in proportion to gas concentration. Greater changes in pulmonary function when SO_2 was administered by tracheal cannula.	Frank and Speizer, 1972
Dogs	22 ppm	30-60 min.	In Vivo	95% of SO_2 was absorbed in upper airways. Portion of non-expired gas was observed in blood stream.	Frank et al., 1967
Swine	35 ppm	Continuously 1-6 weeks	In Vivo	Increased salivation, ocular and nasal irritation, loss of cilia, metaplasia, sneezing and sneezing frequency increased with increase in relative humidity.	Martin and Willoughby, 1971
Beagle dogs	1.0 ppm	1 hour, twice a day for 12 months	In Vivo	Decreased removal of particles by tracheal mucus but no significant changes in pulmonary functions.	Hirsh, 1975
Guinea pigs	1.1 ppm with NaCl, and H_2O mist	few hours	In Vivo	Pulmonary effects with increased humidity.	McJilton et al., 1973

Table I Continued

Type of Animal	SO_2 Concentration	Exposure Time	Type of Test	Results	Reference
Cats	20 ppm with NaCl	few hours	In Vivo	Significant changes in flow resistance.	Corn, 1972
Cynomolgus monkeys	0.1-5.0 ppm with H_2SO_4, and fly ash	few hours/day for 78 months	In Vivo	No synergistic effects with H_2SO_4.	Alarie, 1975
Emphysematous Syrian hamsters	650 ppm	few minutes	In Vivo	Relatively minor influence on airway obstruction.	Goldring, 1970
Influenza infected mice	2.9-19.3	7 days	In Vivo	At low range less pneumonia however at higher range more pneumonia.	Fairchild, 1972, 1975, 1977
Chickens	1.0 ppm with Newcastle virus	few days	In Vivo	Both SO_2 and Newcastle virus slowed clearance rates.	Wakabuyshi, 1977

1. Sulfuric acid has two distinct toxic actions: a) it promotes larynx spasms and bronchispasms (bronchial constriction), and b) it can also produce irreversible scarring of the bronchioles and alveoli.

2. Not all sulfates are irritants. The irritant potency is not related to the sulfate ion as such. Of the compounds tested, if particle size remains the same, the order of irritant toxicity would be: sulfuric acid, zinc ammonium sulfate, zinc sulfate and ammonium sulfate.

3. The particle size of the aerosol is a critical factor in determining both the nature and degree of irritant response. For instance, when sulfuric acid mist and zinc ammonium sulfate were administered to guinea pigs at the same concentrations with nearly equivalent particle size (approximately 0.8μ) sulfuric acid was twice as potent as zinc ammonium sulfate. However, at the same concentration and when the particle size of the zinc ammonium sulfate was 0.3μ and the sulfuric acid remained at 0.8μ zinc ammonium sulfate was more potent by a factor of nearly 3 to 1.

(We note that 80 to 90% of ambient sulfates have been found to be less than 2.0μ in diameter (USHEW, 1969a).)

4. Sulfur produces a less irritant effect if it is present as sulfur dioxide gas than if it is present as particulate sulfate or sulfuric acid. Under laboratory conditions, if SO_2 is converted, depending on size, type of sulfur compound, and degree of conversion, there can be up to a 20-fold increase in toxicity. In ambient air, assuming only a 10% conversion rate of

SO_2 to irritant sulfate of 0.3 μ particle size, Amdur predicts a 4-fold increase of irritant effect (Amdur et al., 1969, 1971).

5. Guinea Pigs appear to be more sensitive than other animal species.

IV. HEALTH EFFECTS EPIDEMIOLOGICAL STUDIES

In studying the impact of air pollution on human health, we must recognize the limitations of applying the results of laboratory studies to the "everyday" human condition. While toxicologic studies on animals are valuable because these permit the careful control of the most important variables - the use of a wide range of exposures, and the examination of body tissue - there are species reaction differences between animals and humans, particularly with regard to the respiratory tract. Analogously, experimental or laboratory exposure of human volunteers allows control of variables, but there are obvious ethical limitations: 1) on the whole,young healthy adults must be used for subjects and therefore results cannot easily be extrapolated to a heterogeneous population; 2) high doses cannot be ethically used with humans, thus there are no experiments with a wide range of exposure levels or experiments with chronic exposures. Moreover, additive and synergistic pollutant interactions that occur in atmospheric chemistry are not generally duplicated in laboratory conditions, hence misleading data on dose-response.

Epidemiologic studies, however, have the advantage of being carried out in the real environment, and much information has been learned from them about the acute and chronic health effects of sulfur oxides (Goldstein, 1975; Neal, 1977). Multiple regression studies, a common epidemiologic approach, attempt to find a correlation between certain environmental factors and certain adverse health effects. The health effects of air pollution, for instance, have been observed by examining statistical relationships between

air measurements and medical data, or by comparing one community with another, or by studying the same community at different exposures.

A. Acute Air Pollution Episodes

The episodes of acute air pollution which have occurred in the Meuse Valley, Belgium, 1930; Donora, Pennsylvania, 1948; London, England, 1952; and 1962; New York City, New York, 1953 and 1963; and in Chicago, Illinois, 1969 provide the strongest evidence of the effect of air pollution on health. In practically all of these acute episodes there resulted significant increases in mortality and morbidity. In the 1952 London episode alone over 4,000 excess deaths were reported.

Air pollution levels during the earlier acute episodes are not fully available. However, we do know that SO_2 levels reached 1.34 ppm in London during the 1952 episode (Logan, 1953; (Morbidity, 1954). In New York City in November, 1953, from 17 to 26 excess deaths per day were reported when a stagnant air mass engulfed the city and SO_2 levels rose to an average of 0.15 to 0.2 ppm (Greenburg et al., 1967). In Chicago during a thermal inversion in November, 1969, daily city-wide levels averaged .071 ppm and the highest city-wide hourly average was 0.295 ppm. In the high pollution community the hourly level reached .412 ppm. Excess deaths from some cardiac and respiratory diseases were attributable to pollution in this episode (Carnow and Namekata, 1977).

In all of these episodes, the very young and the very old experienced more respiratory and heart dysfunctions than other age groups, and their responses were more severe. Those chronically ill, particularly with cardiovascular disease (affecting the heart and blood vessels)and respiratory disease were the most seriously harmed. The mortality rate was much higher in these groups than any other. During the episodes, generally, the deaths from cardiovascular disease occurred early and dropped off sharply, while deaths from pulmonary disease usually began to occur on the second or third day and continued for a longer period of time (Carnow et al., 1966).

Amdur (1969) has noted that during most of these episodes, three factors were usually present: 1) cold temperatures, which increased the solubility of irritant gases in liquids; 2) fog, which provided droplets allowing for the conversion of SO_2 into sulfuric acid mist; and 3) a temperature inversion, which produced a stagnant air mass containing high concentrations of air pollutants. Weather variables, therefore, have a strong impact on the chemical processes and health effects of pollutants like SO_2.

B. Morbidity Studies of Lower Levels of Air Pollution Over Long Periods

1. The CHESS Studies

The most extensive, relatively recent studies on the epidemiologic association of SO_2, sulfates and human health were conducted by the Environmental Protection Agency's Community Health and Environmental Surveillance System (CHESS) for 1970-71, published in 1974.

These studies were made in the early 1970's after pollution levels had already decreased from the levels of the 50's and 60's. However, the report shows adverse health effects from air pollution even at the lowered level. These adverse effects were more consistently associated with exposure to suspended sulfates than to sulfur dioxide or total suspended particulates. Generally the results were:

1. Chronic Bronchitis Rates - In four of the regions, CHESS reported a consistent, statistically significant pattern of chronic bronchitis among residents of the more polluted communities. Smoking contributed more than air pollution to the rates of chronic respiratory disease, and there was considerable variation from one community to another. The

contribution of occupational exposures to chronic respiratory disease was also somewhat larger than that of air pollution, being one-half as large as cigarette smoking. The effects of smoking, industrial exposure, and air pollution appear to be additive. The report concluded that excess bronchitis rates are associated with SO_x exposures alone, at levels of 92 to 95 $\mu g/m^3$ (.03 ppm) SO_2 and 15 $\mu g/m^3$ suspended sulfates.

2. Lower Respiratory Disease Rates - In the Salt Lake Basin and the Rocky Mountain CHESS communities, rates of lower respiratory disease (LRD) were greater among children ages 0 to 12 who had lived in polluted communities for three or more years. The report included that an "excess of respiratory disease (among children 0-12) may reasonably be associated with community exposures of approximately 95 $\mu g/m^3$ (.3 ppm) SO_2 and 15 $\mu g/m^3$ SS (suspended sulfates)."

3. Acute Respiratory Disease Rates - The report determined that "a conservative estimate" would be that exposures to 210 $\mu g/m^3$ (.07 ppm) SO_2, with 104 $\mu g/m^3$ total suspended particulates (TSP) and 16 $\mu g/m^3$ SS were associated with a 5 to 20 percent excess of acute respiratory disease. In New York the study found an association between air pollution and susceptibility to Hong Kong-type influenza among otherwise healthy families.

4. Pulmonary Function Tests - Pulmonary function studies of elementary school children in New York and Cincinnati demonstrated that forced expiratory volume ($FEV_{0.75}$) was diminished by exposure to air pollution.

5. Asthma Attack Rates - Temperature changes were a stronger determinant

of asthma attack rates than any particular pollutant. No relationship with SO_2 and asthma attach rates was found at any temperature. Suspended sulfate levels demonstrated the only consistent relationship with daily aggravation of asthma and cardiopulmonary symptoms. The "best judgment" of the authors was that suspended sulfate exposure as low as 8-10 $\mu g/m^3$ for 24 hours could be a contributory factor to "significant aggravation of pulmonary symptoms" (Health, 1974).

There are certain methodological weakness in some of the CHESS studies, However the CHESS results are not to be considered invalid by any means.

2. Chicago Air Pollution Studies

For over a decade, the Chicago Air Pollution Study Group has been examining the health effects of air pollution, particularly to define those individuals in the population most sensitive to air pollutants, and the levels of pollutants at which their health is adversely affected. The results of the earlier studies are consistent with the CHESS results, however, suspended sulfates were not considered. Though the studies were independent, similar methodologies were employed. Moreover, the air pollution mix in the New York area is not unlike that of Chicago.

The Chicago Chronic Bronchipulmonary Disease Registry Study began August, 1966 (Carnow et al, 1969) with a total of 571 bronchitic patients from 16 different facilities. The patients were classified according to the severity of their disease, and each maintained a daily record of acute chest illness. Their records were then correlated with data of the city of Chicago for each square mile of the city for every 15 minutes of the day. Patients' exposures were determined by the estimated level of pollution in the residence and occupation for each 24-hour period.

Illnesses appeared to correlate with levels of sulfur dioxide, with illness increasing at each of 7 levels of SO_2 pollution. At 0.24 ppm for 24 hours, there was more than twice as many acute chest illnesses as when the level was 0.04 ppm. It appeared that when SO_2 was considered as a pollution index in males 55 and over with advanced bronchitis, there was a relationship between levels of pollution and frequency of acute chest illnesses.

Recently, multiple regression analysis of personal air pollution exposure has been completed. Of special interest to this review is the relationship between maximum temperature, windspeed, personal SO_2 exposure and acute illness for all respiratory diseases. The dependent variable was the percent of excess emergency room visits (ERV) for all respiratory diseases combined. Using .03 ppm (annual mean) - SO_2 explained 4.7% of all emergency room visits (ERV) for respiratory conditions using .14 ppm - it explains 66% of all visits. Data from 48 days are included in the calculations.

3. Hamilton Ontario Air Pollution Studies

A Canadian retrospective hospital admissions study of Hamilton, Ontario, a steel-producing city of about 350,000 people, found a "strong relationship between hospital admissions for acute exacerbations among adults with chronic respiratory illness and among children with acute respiratory disease." This study used an air pollution index (API) which included SO_2 and particulate measurements and climatological data.

4. Allegheny County, Pennsylvania Air Pollution Study

Another recent study with a very good data base and methodology by Carpenter and associates investigated the relationship between hospital costs and exposure to air pollution in Allegheny County, Pennsylvania (Carpenter et al., 1977).

After correcting for race, age, sex, smoking habits, neighborhood income and occupation, they found that respiratory and suspect circulatory system disease showed statistically significant increased hospitalization rates ($P < .01$) and lengths of stay for those exposed to higher levels of SO_2 ($> 99.3\ \mu g/m^3$ or .035 ppm) and particulates ($> 115\ \mu g/m^3$) compared to those from neighborhoods meeting air quality standards. Control diseases were not affected by the air pollution index. Using the area's average costs per day for hospitalization, they estimated total increased costs of hospitalization for the 1.6 million persons in Allegheny County, PA, to be $9.8 million for 1972 ($9.1 million for increased hospitalization rates and $0.7 million for increased length of stay) (Carpenter et al., 1977).

C. Mortality Studies

Deaths from air pollution is usally the end of a cumulative process of stresses and insults and, as such, is a much less precise indicator of the adverse effects of air pollution. Two studies of mortality were carried out by our group some years ago and a third more recently. The first divided square miles of Chicago into low, moderate, and high levels of SO_2 and compared deaths from cardiac and respiratory disease in each square mile of the city; then grouping them into these three categories. Excess deaths were found in the highly polluted from both respiratory and cardiac causes. No effort was made to standardize in this study for socio-economic differences. In 1969 an air pollution episode occurred in Chicago. During this ten day episode there were excess deaths in a number of categories, particularly from respiratory diseases in white males and some cardiac disease categories including rheumatic heart disease, hypertension and in older black males ischemic heart disease. Past mortality studies have been summarized by Goldstein (NAS, 1975) and Finklea (1975). Generally, lower levels of air pollution have yielded less consistent results in mortality studies than in morbidity studies.

Two recent studies were carried out by us examining the impact of individual pollutants on morbidity and mortality.

In regards to mortality - TSP - was significantly related to disease while SO_2 was not when chronic exposure was considered.

When a 25 percent reduction in TSP, which is almost equivalent to the percentage reduction in TSP in Chicago for the period 1970-75, was applied to the models developed, the age-adjusted death rate for all non-accidental causes would be decreased by 5.36% (54.65 deaths per 100,000 persons) in Chicago. A decrease in the death rate by cause was estimated to be 8.85% (all heart diseases), 6.42 (ischemic heart disease), 16.95% (other heart disease), 26.16% (emphysema) and 6.47% (other non-accidental causes). The implications are great and attempts are being made to extend the study.

Models developed in daily analysis also imply that there would be possible acute effects of daily air pollution concentrations (both SO_2 and TSP, in addition to their interaction) on daily mortality changes, controlling for weather changes and day-of-week effects. Based on the model for the day of death onset, it is estimated that a 25 percent reduction in daily levels of each pollutant would decrease daily non-accidental deaths by 1.815% (due to SO_2), 2.045% (due to TSP) and 0.867% (due to an interaction between SO_2 and TSP) in the city of Chicago.

Models for heart disease indicate that the number of daily deaths caused by heart disease could be affected by levels of SO_2, TSP and their interaction. Based on the model for the day of death onset, it is estimated that a 25 percent reduction in daily levels of each pollutant would decrease daily deaths from heart disease by 1.717% (due to SO_2), 2.048% (due to TSP) and 0.940% (due to an interaction between SO_2 and TSP) in the city of Chicago.

THE MORBIDITY STUDY

A morbidity study in which linear regression models have been developed to quantitatively estimate the degree of the air pollution contribution to emergency room visits for cardiac and respiratory diseases in two major hospitals in the city of Chicago was also just completed.

According to the results, sulfur dioxide based on patient exposure levels can account for about 13% of the variation of emergency room visits for acute bronchial and lower respiratory infections and about 22% for total cardiac diagnoses.

Table II summarizes other studies on the effects of SO_2 on humans.

Table II Summary of Selected Humans Studies with SO_2 Alone or in Combination with Other Pollutants (sulfates, acid mists, and virus)

Type of Subjects	Compound and Concentration	Exposure Time	Results	References
11 healthy male adults	SO_2 1,5,13 ppm	10-30 min.	At 1 ppm slight change in pulmonary flow resistance in one subject. At 5 ppm change in pulmonary flow resistance in all subjects. At 13 ppm greater change in pulmonary flow rate. No change in pulmonary compliance, tidal volume breathing rate or pulse rate, only the frequency increased at 13 ppm exposure.	Frank, 1961
9 healthy adults	SO_2 0.5, 1.0 and 5.00 ppm	30-60 min.	SO_2 caused a decrease in maximum expiratory flow (MEF) and 50% decrease in vital capacity (VC) at 1 and 5 ppm. With water aerosol a decrease in MEF was also observed at 0.5 ppm SO_2.	Snell and Luchsinger, 1969
Total Study of 25 healthy adults as total		Total Study 4 year sequential		
a. 13 adults	SO_2 1.00 to 1.17 ppm	1 hour	No change in pulmonary function but significant difference in airway resistance (SRAW) in one subject	Lawther, 1975
b. 12 adults	SO_2 1.00 to 1.17 ppm	25 vital capacity breath	Difference in airway resistance	
c. 17 adults	SO_2 2.8 to 3.3 ppm	8 deep breaths from a boy	SRAW increase by a factor of 3	
d. 10 adults	SO_2 30 ppm	10 minutes	There was a wide range of sensitivity. SO_2 by itself could be a contributing factor to change in lung function or exacerbation of bronchitis.	

Table II Continued

Type of Subjects	Compound and Concentration	Exposure Time	Results	References
7 healthy adults	SO_2 16.1 ppm (average)	25-30 minutes for 5 exposures	All SO_2 was absorbed in nasal passages, less coughing, less irritation to throat, fewer and smaller increases in flow resistance than when exposed to same conc. by mouth	Speizer and Frank, 1966
9 healthy non-smoking adults	SO_2 5.0 ppm	20 breaths by mouth	No significant effect on ciliary mucus clearance	Wolff, 1975
15 healthy males	SO_2 1.0,5.0 and 25.0 ppm	6 hours in each concentration	Significant decrease in ciliary mucus flow at 5.0 and 25.0 ppm	Anderson, 1973
5 healthy males	H_2SO_4 mist of particle size 0.35u at conc. of 5.0 mg/m^3	15 minutes	Respiratory rate increased. Maximum expiratory flow rate decreased by 20%, tidal volume decreased by 28%	Amdur, 1952
5 healthy males	H_2SO_4 (1.5 u - 30 u particle size) 3-38 mg/m^3 conc. at 91% relative humidity	183 exposures each of 16 min. by mask & 31 exposures each of 60 min. in chamber	Increase airway resistance 35-100% above normal rate and much increase up to 150% in high humidity	Sim and Pattle, 1957
4 healthy males	London air from 1965 to 1971	2-5 km. walk to work	During first year study one subject showed increase airway resistance with pollution, relative humidity and lower temp. After the 1st year no effects were found.	Lawther, 1977

Table II Continued

Type of Subjects	Compound and Concentration	Exposure Time	Results	References
8 healthy male students	SO_2 0.37 ppm O_3 0.37 ppm separately then combined	2 hour period	More adverse effect due to combination	Bates, 1975
10 human adults	SO_2 5 ppm	4 hours	50% decrease in nasal mucous flow rates	Anderson, 1973

SUMMARY

Animal and human toxicologic studies have identified many of the mechanisms of adverse health effects of sulfur oxides. Pure SO_2 is an irritant gas that produces irritation and inflammation of the tissue that it directly contacts. The principle observed effects of acute exposure are: 1) airway resistance, (2) mucociliary impairment, (3) an acute bronchospastic effect , and (4) by the above, interference with breathing. Chronic exposure may cause chronic damage to the respiratory system. Acid sulfates are also classified as irritants and are more potent than SO_2. The health effects of other sulfates, sulfites and bisulfites levels not yet been determined, but there is some evidence they may be mutagenic (Hickey et al., 1976). The known target organs in humans for SO_2 and irritant sulfates are principally the lungs and secondarily the heart because of its high oxygen need and its interrelationship with the respiratory system.

The best available information on the acute and chronic adverse health effects of sulfur oxides comes from epidemiologic studies carried out in the real environment. Some of these studies have shown adverse health effects at below ambient standards. At present epidemiologic studies have observed associations of SO_2 exposure and other pollutants with:

1. increased mortality from cardiac and respiratory disease (Carnow and Namekata, 1977; Goldstein, 1975; Rall, 1974).
2. increased sputum cellularity, in healthy adults, indicating inflamed lung tissue (Nobutomo, 1978).
3. increased incidence of chronic respiratory disease (asthma, bronchitis, emphysema) and possibly cancer (Goldstein, 1975).
4. increased incidence of acute respiratory attacks among those at high risk because of chronic pulmonary disease (Carnow, 1969).

5. increased incidence of cardiac death among those at high risk because of cardiac disease (Carnow, 1969).
6. increased rate of hospitalization and increased length of hospital stay (Carpenter, 1977)
7. increased emergency room admissions on days of high pollution (Carnow and Namekata, 1977)
8. increased loss of work days due to respiratory distress (CHESS, 1974).
9. increased absence from school due to respiratory distress (CHESS, 1974).

Those people at high risk include infants and children, male adults, aged 55 years and over with severe (advanced) chronic bronchitis, smokers, and "marginal" people with poor adaptive capacity, such as those suffering from chronic lung disease, heart disease, asthma, and certain congenital diseases.

Laboratory tests have generally used much higher concentrations of SO_2 to produce adverse response than have epidemiologic studies. These higher levels are required in laboratory tests because: 1) They often use SO_2 by itself, whereas the ambient air SO_2 converts to the more toxic sulfuric acid and often sulfur products, and in the presence of high humidity and/or other pollutants this conversion may be relatively rapid. 2) The test use animal species and strains of species that may be more resistant to SO_2 than humans. 3) When humans are used, they are usually normal young adults.

Studies are needed to further quantitate the impact of SO_2 as acid sulfate and its contribution to morbidity and mortality as a part of a mix of pollutants and meterologic variables.

While the available data base is not yet sufficient for precise dose/effect functions, a recent Chicago air pollution study allows a projected linear relationship between acute morbidity and personal exposure to SO_2 based on actural results. This, together with previous estimates, (Fishelson and Graves, 1977) can assist policy makers in decision making.

CONCLUSIONS AND RECOMMENDATIONS

1. There is not health justification for relaxing present SO_2 ambient standards, nor is there likely to be. Moreover, present SO_2 standards may not be protecting high risk groups and hypersensitive individuals.

2. While the data base on sulfates from epidemiologic studies is insufficient, animal and human toxicologic studies have shown that many sulfates are unquestionably considerably more toxic than sulfur dioxide. There is also no question that there is a build-up of sulfates in the air of certain regions of the U.S. notably the Northeast. It also appears that sulfur dioxide levels are not good measure of ambient sulfates. For those reasons it is clear that a separate sulfate standard is desirable to protect health. Since adverse health effects were observed at levels of 8 to 10 $\mu g/m^3$ it would be prudent to promulgate a temporary standard, possibly at half of that or of 4 $\mu g/m^3$ annual mean, not to be exceeded more than one percent of the time during the course of a year.

3. Reductions of ozone and particulate which appear, to act synergestically with SO_2 might result in reduced negative health effects of all 3 pollutants.

4. Additional toxicologic and epidemiologic research on possible mutagenic, carcinogenic and cocarcinogenic effects of SO_2, sulfates, sulfites and bisulfites is indicated.

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ENERGY, ENVIRONMENTAL, AND ECONOMIC IMPACTS OF FLUE GAS DESULFURIZATION UNDER ALTERNATIVE NEW SOURCE PERFORMANCE STANDARDS

by

Andrew J. Van Horn, Richard A. Chapman,
George Ferrell, Peter M. Cukor, and
David B. Large

Teknekron, Inc.
2118 Milvia Street
Berkeley, CA 94704

ABSTRACT

The energy, environmental, and economic impacts of flue gas desulfurization (FGD) under alternative revisions to the New Source Performance Standards for coal-fired utility boilers have been examined using the Utility Simulation Model (USM). The USM simulates investment and operating decisions related to choices of fuels and pollution control equipment through the use of extensive data bases, and cost and performance models for pollution control, including control of SO_2, particulates, and NO_x. For each of the 48 contiguous states in the U.S., alternative projections have been made of the impacts of utility operations from 1976 to beyond the year 2000.

The SO_2 control technology and cost model is structured to calculate capital costs, variable operating costs, and capacity penalties for limestone, lime, and magnesium oxide FGD systems installed in module sizes of between 50 and 130 MW each, except for systems of less than 50 MW. All systems of 100 MW or more include a spare module for added reliability. The model is applied on a unit-by-unit basis utilizing a data base with detailed information on each existing and announced coal-fired utility boiler in the U.S.

In calculating FGD system costs and penalties, the model considers both the gas flow rate and the quantity of SO_2 to be removed. This level of sophistication makes it possible, for instance, to compare FGD costs for the same coal at various emission limits, or for various coals with the same sulfur content but different heating values. This model is part of both the Coal Assignment Model and the USM planning module and is used in selecting a fuel and pollution control strategy as well as calculating the operating and cost functions for individual generating units in each year of the simulation.

In this paper we summarize the recent sensitivity studies performed for EPA concerning the revised New Source Performance Standards to be established for coal-fired electric utility boilers.

ENERGY, ENVIRONMENTAL, AND ECONOMIC IMPACTS OF FLUE GAS DESULFURIZATION UNDER ALTERNATIVE NEW SOURCE PERFORMANCE STANDARDS

INTRODUCTION

Over the past three years, under the auspices of the U.S. Environmental Protection Agency, Teknekron has developed and applied its Utility Simulation Model (USM) to examine a variety of energy and environmental problems. For the past 18 months we have used the USM to review the economic and environmental impacts of various revised New Source Performance Standards (NSPS) for SO_2 and particulates from coal-fired electric utility boilers.

This paper contains a summary of the recent results of our NSPS Phase Three analyses,(1) which focused on critical uncertainties surrounding a number of key factors that will influence future impacts of the revised NSPS. These factors will affect utility costs and hence will influence the coal choices and pollution control measures adopted by utilities in response to alternative standards. Teknekron has carried out city-specific analyses of utility coal and pollution control choices and their sensitivity to the factors of interest. Complementing these sensitivity analyses are our state, regional, and national impact projections for alternative standards for the period from 1976 to the year 2000.

Key elements we have varied include coal mine prices, coal transportation rates, coal sulfur and Btu contents, and the costs and performance of FGD scrubbers. In each case, the selected range of variation reflects the element's degree of uncertainty and sensitivity to critical issues.

The implications of these variations for the Environmental Protection Agency's (or anyone else's) ability to distinguish between similar standards are discussed. Also discussed are the sensitivities of several cost-effectiveness calculations (for example, cost per ton of SO_2 removed) which have been posited as measures of the worth of various standards.

The impacts of revised standards will depend not only on utility coal and pollution control choices but also on such factors as the future growth in electricity demand, the amount of nuclear capacity, the phasing out of gas steam plants, and the price of oil. These factors are themselves subject to uncertainty. In our projections for 1976 to 2000, we have used the latest assumptions made for these parameters by the joint EPA/DOE working group.(2)

UTILITY SIMULATION MODEL DESCRIPTION

The Utility Simulation Model consists of a number of interconnecting computer modules and data bases that simulate decisions for system planning and operation, utility finance, and the operation of individual technical processes. The model is driven by a set of exogenous scenario elements that include electricity demand levels, financial market conditions, fuel prices and availabilities, advanced technology deployment, and environmental regulations. For each scenario, the model calculates the following by geographic region (county or state) for future years up to 2010:

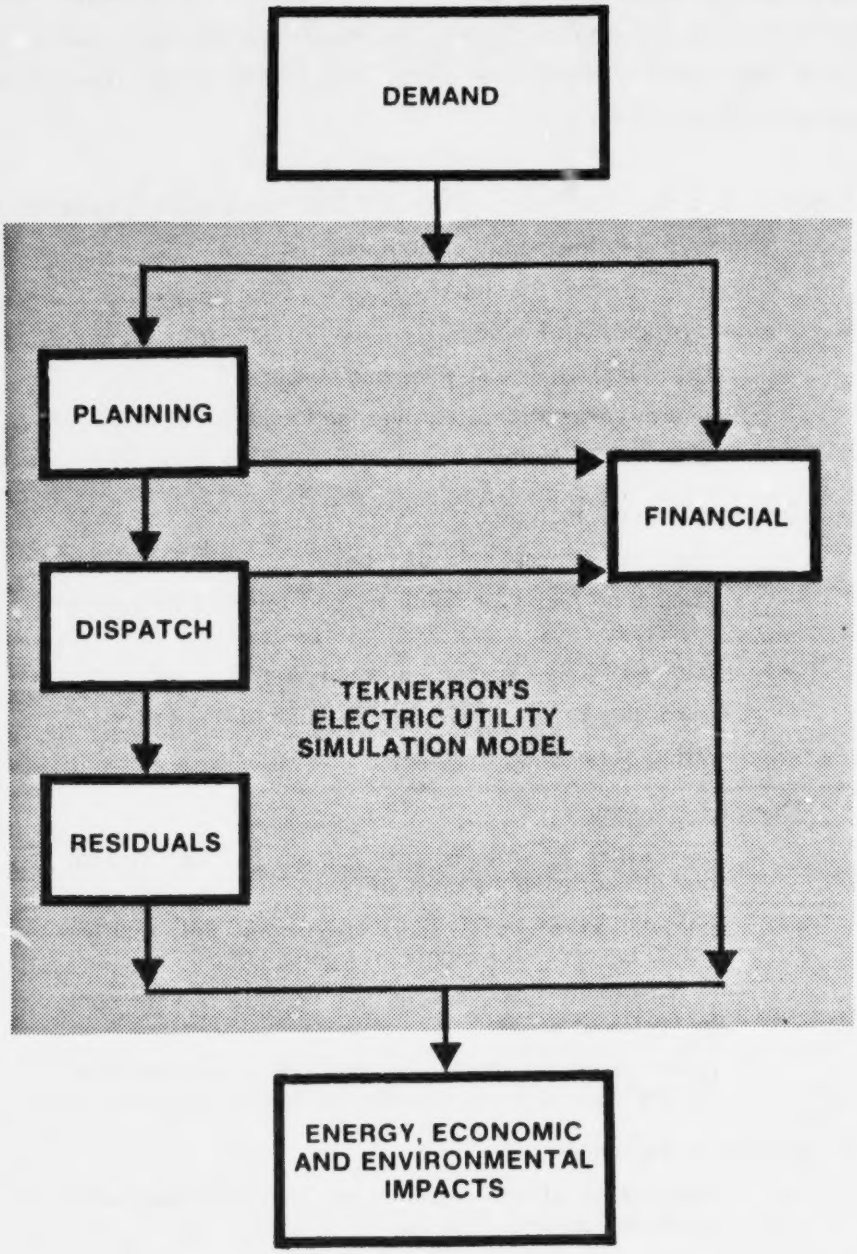
- Factor demands, including
  - fuel use, by type and for coal by region of origin
  - electricity generated
  - Capital requirements, by source (e.g., debt, common equity, preferred equity)
  - plant and equipment requirements
  - releases of air and water pollutants and generation of solid wastes
- Financial statistics for utility firms
- Average electricity prices

In order to produce these calculations at the required level of detail, the model considers generating unit sites located in each county where electricity is produced, fuel and water are consumed, and pollutants are released. Since utilities operate as integrated systems, the model presently simulates joint operation (i.e., dispatching) of all generating units within a state. Finally, the responses of utility firms to the external environment in which they function may be changed by the model user modifying present data bases or specifying alternate choices for future system planning and system operation. For example, the particular scenarios evaluated in the New Source Performance Standards Review encompass a range of futures for electricity demand, fuel selection, choices of technology, and pollution control regulations as specified by the U.S. Environmental Protection Agency.

Figure 1 is a simplified diagram of the Electric Utility Simulation Model. The model includes the following major components:

- Demand projection, including
  - retail and wholesale sales and purchases
  - energy generation, i.e., average load growth
  - peak load growth
- System planning, including
  - choice of generating unit type
  - choice of fuel type, quality, and for coal by region of origin
  - choice of pollution control technology
  - expansion of transmission and distribution networks
  - siting of generating units
- Dispatch, including
  - calculation of unit capacity factors for each typical day of operation, by class of unit
  - calculation of total fuel, operation, and maintenance expenses for electricity generation
  - projection of fuel consumption, by type and for coal by region of origin
  - pollution control costs and operating characteristics for the various types of pollution control devices
- Financial, including
  - integration of projected production expenses with construction expenditures

Figure 1
Electric Utility Simulation Model


DEMAND
PLANNING
FINANCIAL
DISPATCH
TEKNEKRON'S ELECTRIC UTILITY SIMULATION MODEL
RESIDUALS
ENERGY, ECONOMIC AND ENVIRONMENTAL IMPACTS

  - projection of the firm's balance sheet, income statement, sources and uses of funds, and other financial statistics
  - calculation of annual revenue requirements and electricity prices

- Residuals, including
  - projection of release rates at the generating unit site for numerous air and water pollutants and for solid wastes
  - projection of consumption of water and other resources

Teknekron's coal assignment model (CAM) is used to select coals to be used in the USM for each state and regulatory category of coal plant (e.g., SIP, NSPS, RNSPS). A maximum of 50 distinct coals from 12 different supply regions may be considered for use in each of the 48 states.

The criterion for coal selection is least levelized annual cost, where cost is determined on an "as-burned" basis. Since the program is interactive, the user may exercise expertise in coal selection and/or knowledge of historical coal movements to force the selection of a coal other than the least cost coal. To facilitate the selection process, CAM produces a supplementary report which includes complete cost information about the first five choices ranked on a least cost basis.

The determination of "as-burned" cost considers the entire fuel-cycle. The CAM program tracks a candidate coal along the fuel cycle from mine production to combustion in the utility plant boiler including particulate control and flue gas desulfurization. In calculating the total "as-burned" cost, the CAM considers the following fuel cycle component costs:

- FOB mine price
- Cost of coal cleaning (where applicable)
- Transportation and handling charges
- Cost of boiler modification (where applicable)

- Particulate control cost
- Cost of flue gas desulfurization

In addition to the above costs, the user may define a "premium" to be added to the delivered cost of any coal. This premium is a surrogate for institutional factors or other influences upon the marketability of a coal which are either imprecisely defined or otherwise not amenable to economic modeling. The total "as-burned" cost is defined as the sum of the component costs in the fuel cycle.

FGD COST MODEL

Teknekron has developed FGD cost and performance models based on PEDCo and TVA engineering and cost estimates for lime and limestone systems and PEDCo cost estimates for magnesium oxide systems.(3,4,5) The models can be used to predict new or retrofit FGD costs for generating plants of between 25 MW and 2,000 MW in size burning coal of any sulfur content and meeting any emission limit.

The three FGD systems are modular in design, with module sizes of between 50 MW and 130 MW except for plants of less than 50 MW in size. One redundant module is included for all systems of 100 MW or greater for a design reliability of 90 percent. The design of the three FGD systems is based on a three-stage turbulent contact absorber (TCA).

The TVA and PEDCo FGD cost estimates represent a reasonable range of costs for use in our sensitivity studies. The PEDCo costs are higher than TVA's and are probably representative of costs that may be used by utilities without extensive experience with FGD systems. The TVA costs, on the other hand, are less conservative and represent cost estimates that may be used in the future by utilities which have had favorable FGD experience. These two cost estimates may also be viewed as representing two points on the FGD "learning curve" with the lower cost estimates indicative of lower, future FGD costs.

Tables 1 and 2 present PEDCo and TVA capital and operating cost estimates for a 500 MW limestone FGD system designed for 85 percent SO_2 removal from a medium sulfur content coal with a 24 hour averaging time. The total capital investment estimated by TVA is about 60 percent of PEDCo's estimate. Total operating and maintenance costs (less fixed costs) are estimated at about 6.0 and 4.7 million dollars respectively by PEDCo and TVA.

The cost of electricity and steam required to operate the FGD system is not calculated by the FGD cost model; instead, electricity and steam requirements are used to calculate unit capacity penalties and are accounted for in this manner by the Utility Simulation Model. For the case illustrated in Tables 1 and 2, the TVA capacity penalty is 2.96 percent, and the PEDCo capacity penalty is 4.25 percent. Either of these estimates is reasonable and representitive of the range of capacity penalties which might be expected.

Within the model, plant characteristics, coal properties, and emission limits are used to determine the required rate of sulfur dioxide removal in pounds per hour and the required gas flow rate in actual cubic feet per minute for an FGD system having an annual average removal efficiency of 90 percent or greater. If a given generating plant needs to remove less than 90 percent of the SO_2 produced to meet applicable emission limits, an FGD system with an efficiency of 90 percent will be used to scrub a portion of the flue gas. The remaining flue gas will be bypassed and mixed with the scrubbed gas to yield the required SO2 emissions and to reduce or eliminate the fuel required for reheating the flue gas. If 90 percent or more of the SO_2 must be removed, an FGD system having the required efficiency up to the limits of technology will be used to scrub the entire flue gas stream.

The cost of such equipment as pumps, hold tanks, feed preparation equipment, and sludge ponds is based on the sulfur dioxide removal rate, while the cost of such items as fans, absorbers, and soot blowers is based on the gas flow rate. Likewise, operating costs are based on either the sulfur dioxide removal rate (e.g., raw material) or the gas flow rate (e.g., electricity, reheat steam or oil).

Table 1. Comparison of TVA and PEDCo Limestone FGD Capital Costs*

Capital Cost Item	PEDCo**	TVA**
Direct costs		
Limestone preparation	$ 2,423,800	$ 3,322,100
SO_2 scrubber	21,012,600	14,786,800
Sludge disposal	1,201,900	2,248,900
Sludge pond	5,632,800	0***
Raw material inventory	162,600	0
Total direct costs	$30,433,700	$20,357,800
Indirect costs	9,271,900	7,348,700
Contingency and fee	10,283,000	3,053,700
Total capital investment	$49,988,600	$30,760,200

* Basis: Coal sulfur content = 2.50 lbs S/10^6 Btu
Sulfur RSD = 0.15, no exemptions
Design sulfur content = 3.63 lbs S/10^6 Btu
Plant size = 500 MW
Five scrubber modules at 125 MW each
85 percent 24-hour average SO_2 removal
1975 costs and dollars

** Costs predicted by Teknekron's SO_2 control model. Not included are interest during construction, working capital, and taxes; these are calculated in the Utility Simulation Model's financial module.

*** Sludge pond capitalization included in sludge disposal operating cost (see Table 2).

Note: More recent estimates by TVA include about $7 million for the sludge pond and a contingency and fee of 25 percent of total direct costs. Total TVA investment estimate is therefore increased to about $42 million.

Table 2. Comparison of TVA and PEDCo Limestone FGD Operating Costs*

Cost Item	PEDCo**	TVA**
Limestone	$ 804,400	$ 769,900
Labor	406,500	783,400
Maintenance	3,736,600	1,816,800
Water	38,000	21,800
Sludge disposal	996,100	1,219,700
Analysis cost	0	69,400
Total O&M costs	$5,981,600	$4,684,000

* Basis: Coal sulfur content = 2.50 lbs S/10^6 Btu
Plant size = 500 MW
85 percent 24-hour average SO_2 removal
Capacity factor = 0.65
1975 costs and dollars

** Costs predicted by Teknekron's SO_2 control model. Not included are: (a) steam and electricity costs, which are used in the Utility Simulation Model to calculate capacity penalties; and (b) fixed charges, which are calculated in the Utility Simulation Model's financial module.

Note: More recent estimates by TVA include a higher cost for maintenance (due to higher capital cost) and sludge disposal. Total TVA operating cost estimate is now about the same as the PEDCo estimate.

Outputs from the FGD model include:

- Capital cost
- Fixed operating cost (independent of plant capacity factor)
- Variable operating cost (dependent on capacity factor)
- Removal efficiency
- Scrubber size
- Capacity penalty (plant capacity used to operate the FGD system)
- Heat rate penalty (accounts for fuel required to operate the FGD system)
- Water used and water cost
- Oil used for magnesium oxide regeneration
- Oil used for reheat
- Annual sludge generation

SO_2 emissions are calculated on the basis of the uncontrolled emission rate and the required removal efficiency.

Input data required for the FGD model include:

- Individual generating-unit characteristics
  - size
  - age (new or retrofit)
  - heat rate
- Coal properties
  - heating value
  - composition (C, H, O, N, S, H_2O, ash)
  - class (bituminous, subbituminous, lignite)
- Environmental factors

  - emission limit (specific limits: percentage removal, ceiling, floor, and averaging time)
- Economic factors
  - year scrubber was built (escalation, inflation)

SENSITIVITY STUDIES

The Coal Assignment Model and the Utility Simulation Model were used to determine the sensitivity of future impacts of various revised New Source Performance Standards to uncertainties in future coal prices, transportation rates, coal properties, and FGD costs.

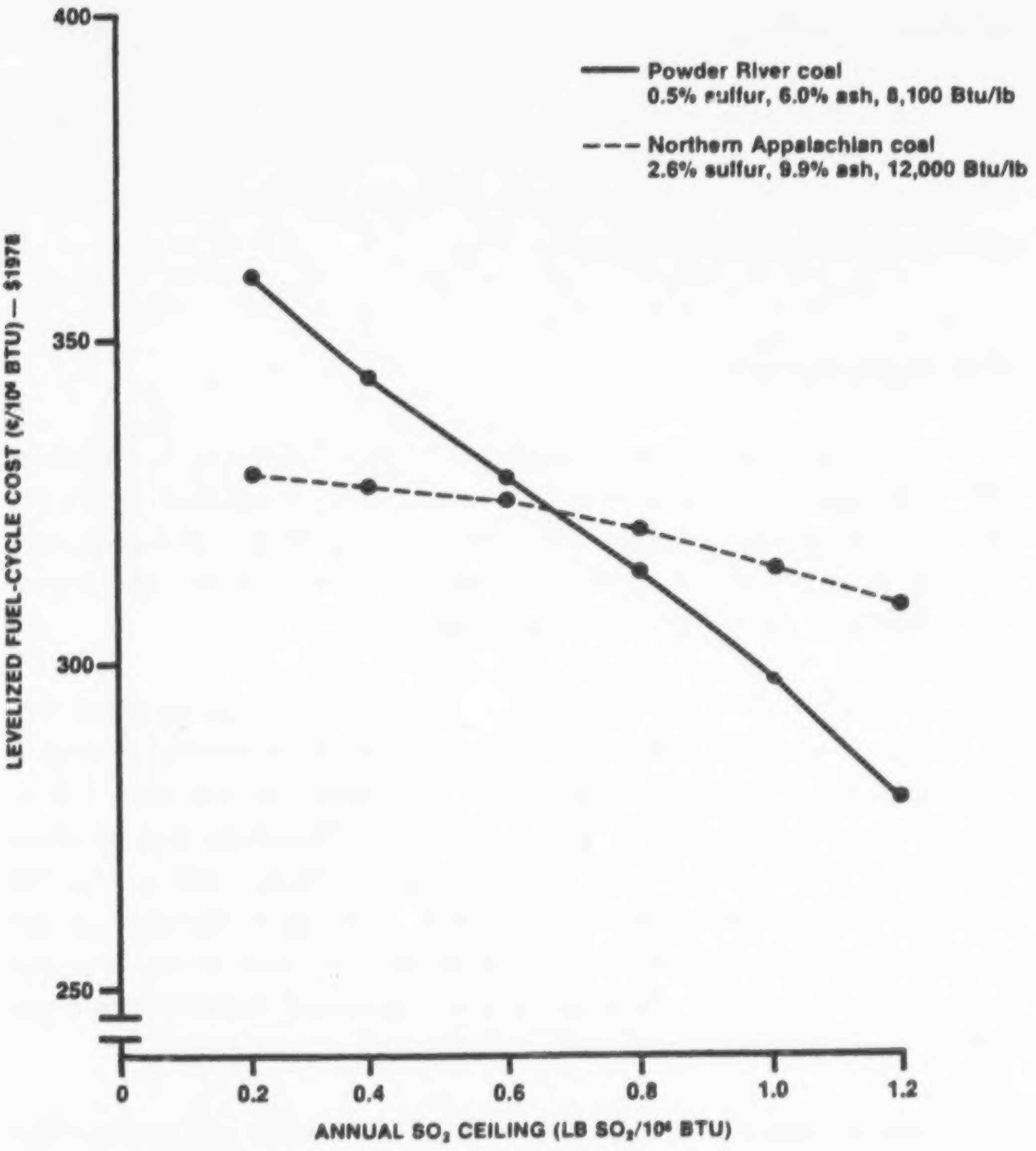
City Specific Analyses

City specific analyses were conducted for key locations (e.g., Columbus, Ohio; Indianapolis, Indiana; Orlando, Florida; and Austin, Texas) to determine the effect of various SO_2 standards on the levelized fuel-cycle cost for various coals and the sensitivity of these effects to uncertainties in future coal prices, transportation rates, coal properties, and FGD costs.

The sensitivity of levelized fuel-cycle cost with respect to the annual SO_2 ceiling for Powder River coal and Northern Appalachian coal used by a utility in Columbus, Ohio is illustrated in Figure 2. For ceilings greater than about 0.65 lb $SO_2/10^6$ Btu, Powder River coal is less expensive and for ceilings less than 0.65, Northern Appalachian coal is less costly to use. Figures 3 and 4 show the sensitivity of the levelized fuel-cycle cost with respect to 24-hour SO_2 floor and respectively transportation rate and FOB mine price. As would be expected, the fuel-cycle cost of Powder River coal is more sensitive to transportation rates, and Northern Appalachian coal is more sensitive to FOB mine price.

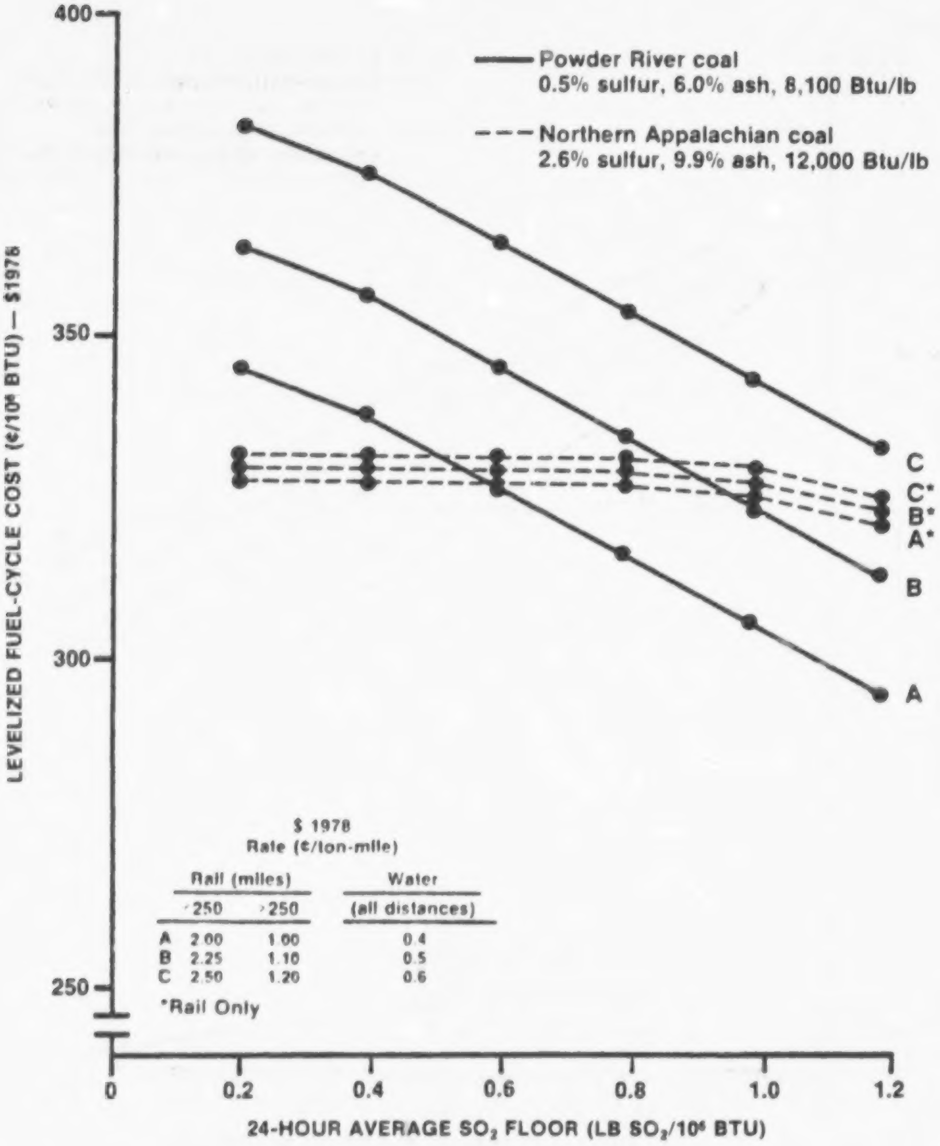
Another important consideration is the sulfur content and heating value assumed for Powder River coal. Figure 5 illustrates the sensitivity of levelized fuel-cycle cost with respect to various 24-hour SO_2 floors and coal characteristics typical for Powder River coals. These curves represent coals with sulfur

Figure 2
Sensitivity of Levelized Fuel-Cycle Cost with Respect to Annual SO2 Ceiling
(Columbus, Ohio)



a) Assumes no mandatory percentage removal requirement

Figure 3
Sensitivity of Levelized Fuel-Cycle Cost with Respect to 24-Hour SO2 Floor and Transportation Rate (Columbus, Ohio)



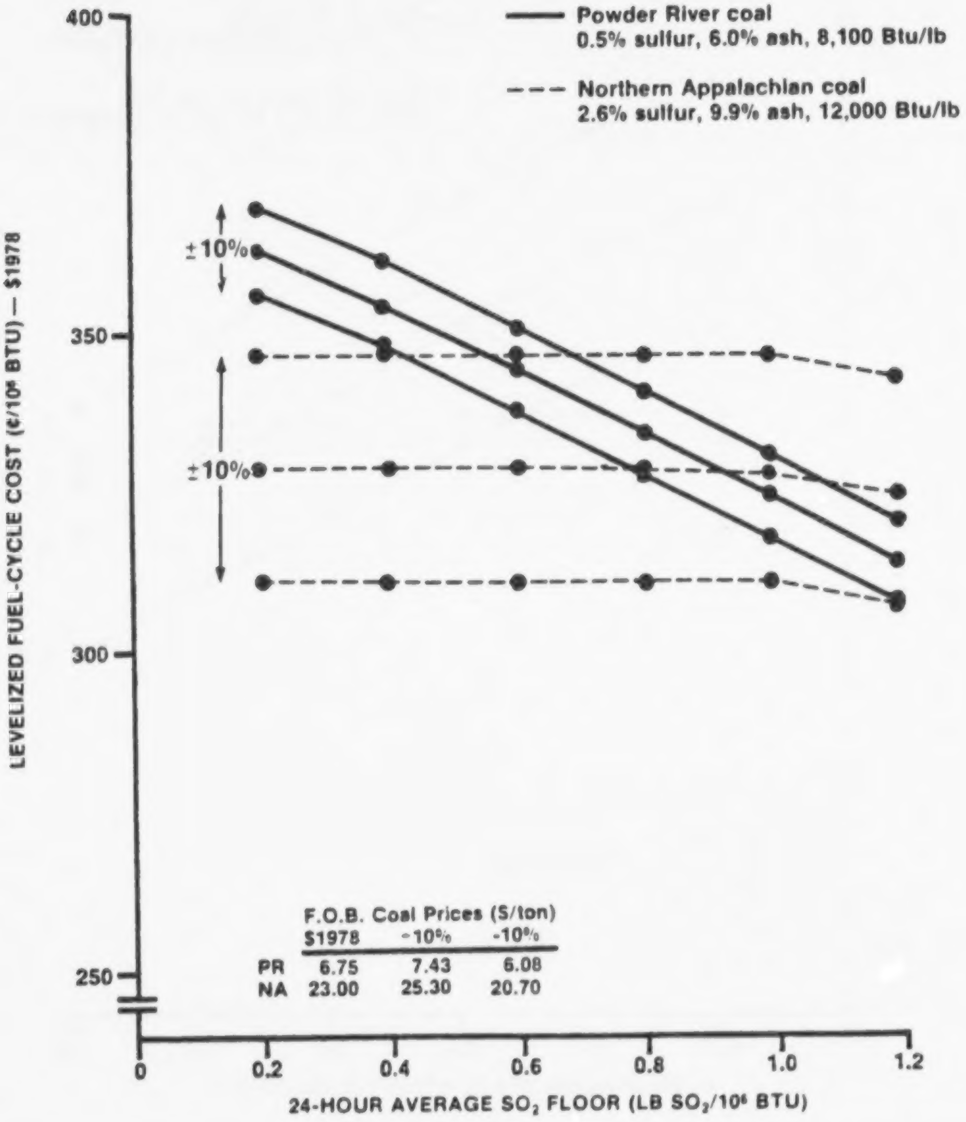
$ 1978
Rate (¢/ton-mile)

	Rail (miles)		Water
	<250	>250	(all distances)
A	2.00	1.00	0.4
B	2.25	1.10	0.5
C	2.50	1.20	0.6

*Rail Only

a) Assumes 1.2 lb $SO_2/10^6$ BTU ceiling with 85% removal, 24-hour average with three day per month exemptions.

Figure 4
Sensitivity of Levalized Fuel-Cycle Cost with Respect to 24-Hour SO2 Floor and F.O.B. Coal Mine Prices (Columbus, Ohio)



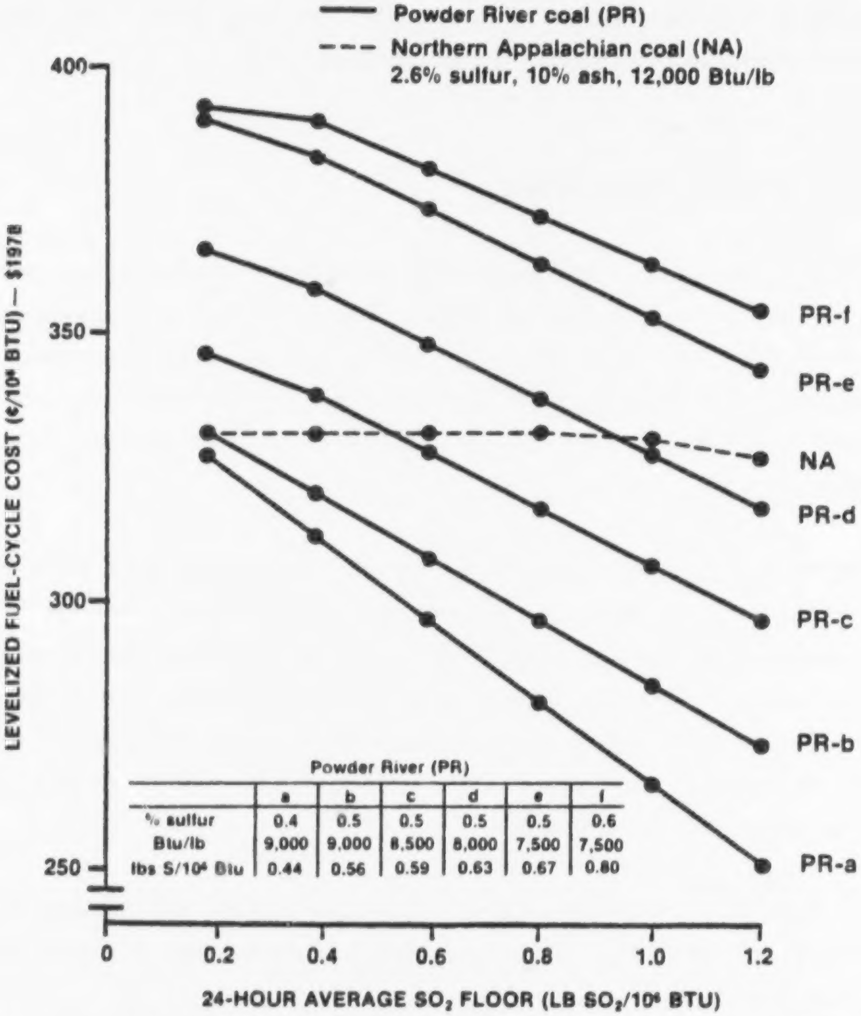
F.O.B. Coal Prices ($/ton)

	$1978	-10%	-10%
PR	6.75	7.43	6.08
NA	23.00	25.30	20.70

a) Assumes 1.2 lb $SO_2/10^6$ BTU ceiling with 85% removal, 24-hour average with three day per month exemptions.

b) Transportation Rates: Rail < 250 miles, 2.25¢/ton-mile; > 250 miles, 1.20¢/ton-mile; Water 0.5¢/ton-mile

Figure 5
Sensitivity of Levelized Fuel-Cycle Cost with Respect to 24-Hour SO2 Floor and Powder River Coal Characteristics (Columbus, Ohio)



Powder River (PR)

	a	b	c	d	e	f
% sulfur	0.4	0.5	0.5	0.5	0.5	0.6
Btu/lb	9,000	9,000	8,500	8,000	7,500	7,500
lbs S/10⁶ Btu	0.44	0.56	0.59	0.63	0.67	0.80

a) Assumes 1.2 lb $SO_2/10^6$ BTU ceiling with 85% removal, 24-hour average with three day per month exemptions.

b) Powder River $1978/ton = 6.75; Northern Appalachia $1978/ton = 23.00

contents between 0.44 and 0.80 lb S/10^6 Btu and are representative of the coals available in the Powder River Basin.

The estimated cost of FGD systems reflects perhaps the greatest uncertainty and as illustrated in Figure 6 is of utmost importance in the selection of the lowest levelized fuel-cycle cost strategy to meet various revised New Source Performance Standards. If the higher PEDCo FGD cost estimates are used, coal selection is affected by the emission limit. On the other hand, if the TVA costs are used, the local coal will be selected for all emission limits.

National Utility Simulation Model Results

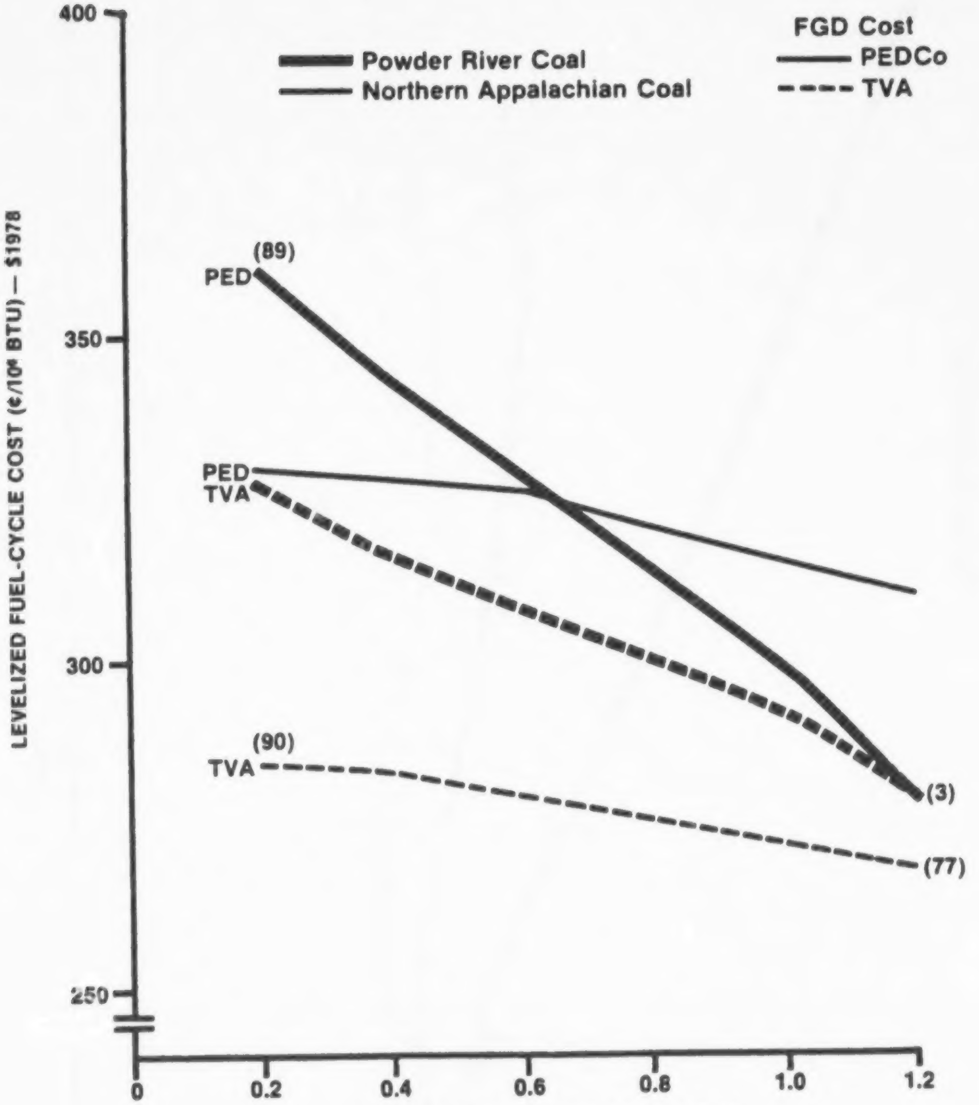
The Utility Simulation Model was used to evaluate the economic and environmental impacts of alternative revised New Source Performance Standards for coal-fired electric utility boilers. Numerous full and partial scrubbing scenarios were evaluated and compared to the current NSPS. PEDCo and TVA FGD cost estimates were used to represent the likely range of uncertainty in total fuel-cycle costs.

The impact of various full and partial scrubbing scenarios using PEDCo FGD costs on national SO_2 emissions through the year 2000 is illustrated in Figure 7. National emissions in the partial scrubbing senarios begin to increase again between 2000 and 2010 while the full scrubbing SO_2 emissions do not begin to increase until sometime after 2010. National emissions by various classes of coal-fired plants in 1995 are presented in Figures 8 and 9 for high (PEDCo) and low (TVA) fuel-cycle costs respectively.

The national percent increase in total utility cost and percent decrease in SO_2 emissions in 1995 for alternative revised NSPS are presented in Figure 10 for the high fuel-cycle cost case.

Utility coal production and Western coal shipments in 1995 for the high-fuel cycle cost (PEDCo) and the low fuel-cycle cost (TVA) cases as a function of

Figure 6
Sensitivity of Levelized Fuel-Cycle Cost with Respect to FGD Cost (Columbus, Ohio)



( ) = Percentage SO_2 Removal

Figure 7
National Power-Plant SO_2 Emissions
(10^6 Tons)
Pedco FGD Costs

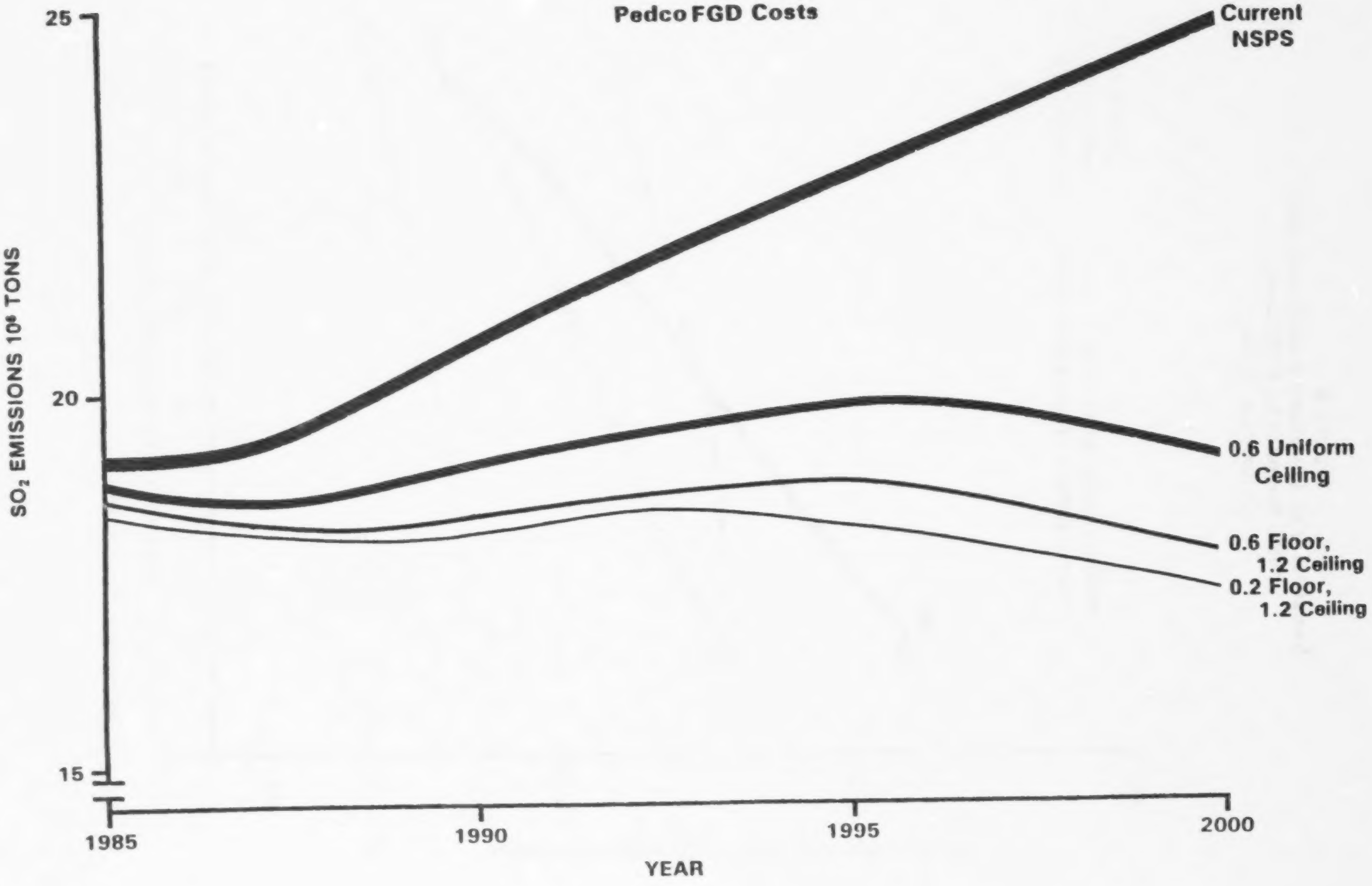


Figure 8
National SO_2 Emissions from Coal Fired Power Plants
(10^6 tons) 1995
Pedco FGD Costs

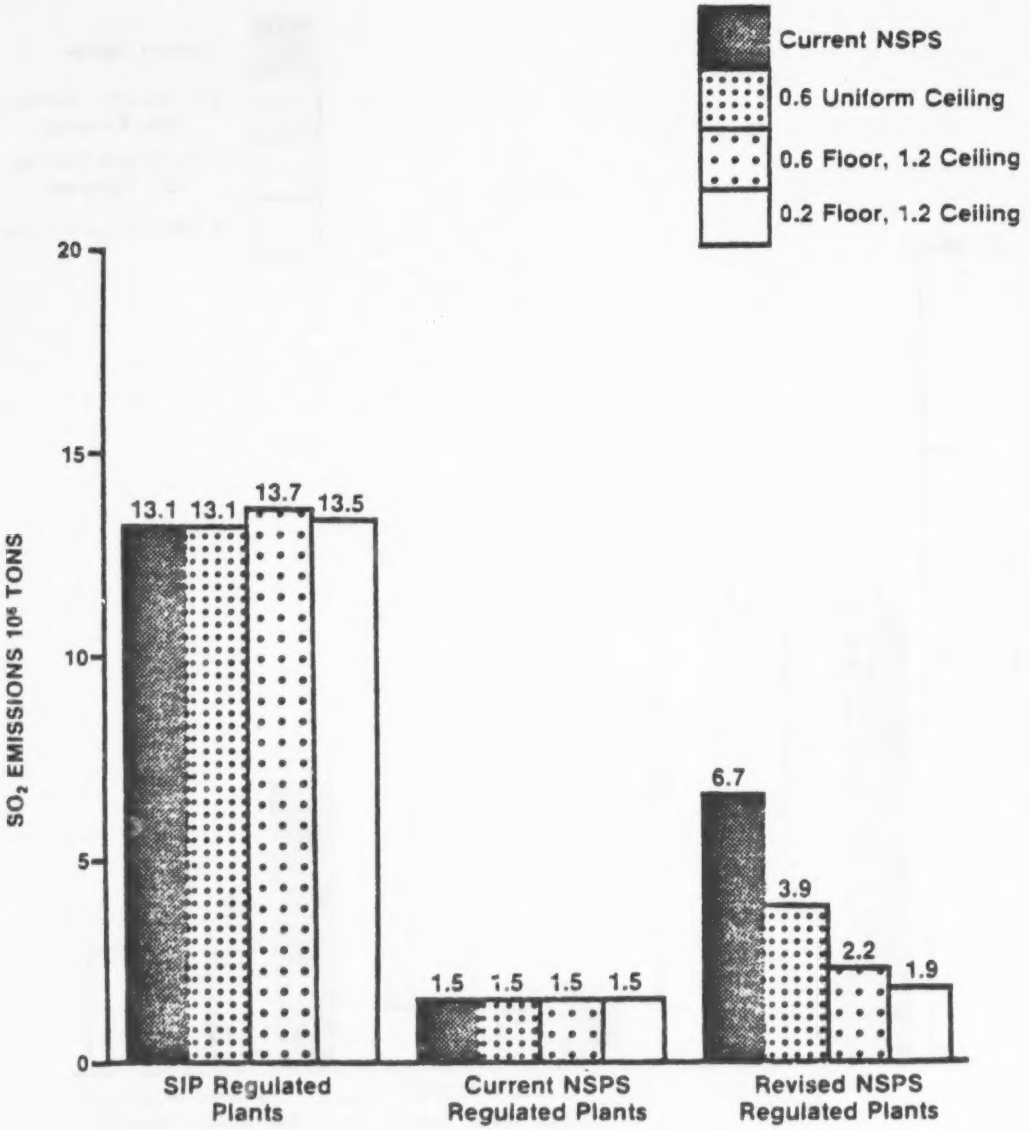


Figure 9
National SO_2 Emissions from Coal-Fired Power Plants
(10^6 tons) 1995

TVA FGD Costs

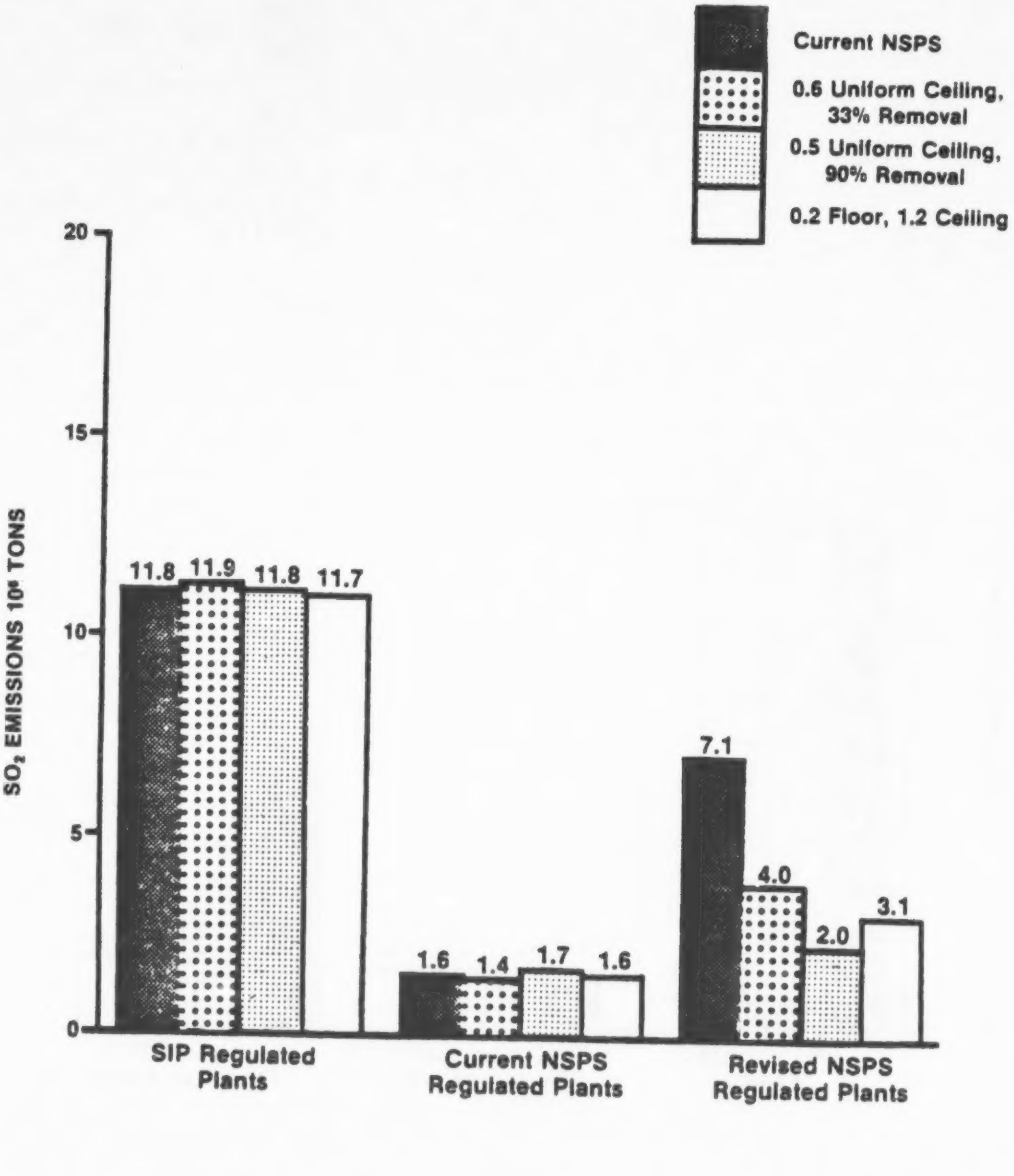
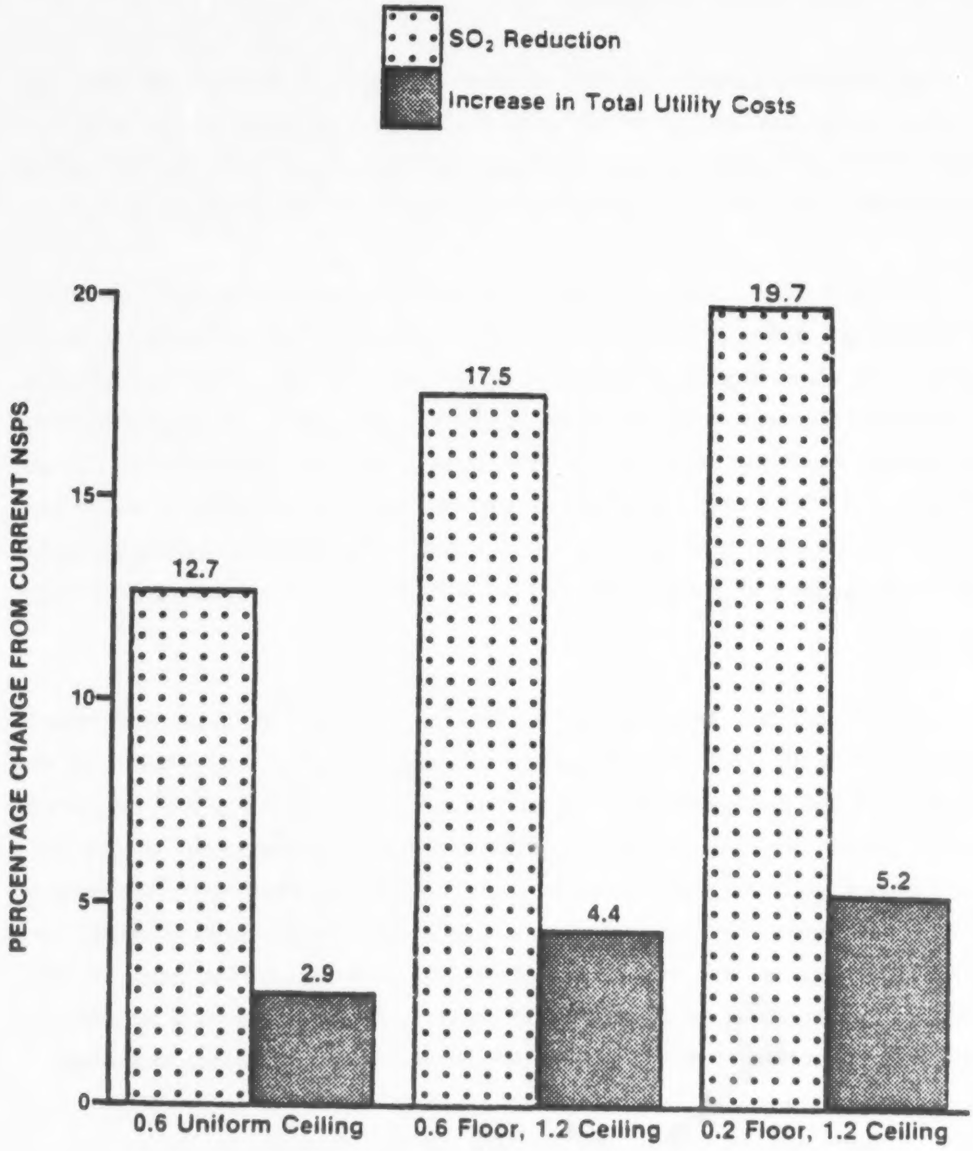


Figure 10
National Percentage Increase in Total Utility Cost and Percentage Decrease in SO_2 Emissions for Revised NSPS
1995

Pedco FGD Costs



the revised New Source Performance Standard are illustrated in Figures 11–14. These figures show the great sensitivity of regional coal production and shipments to relatively small changes in total fuel-cycle costs and the lower sensitivity to the standard itself. It is also clear from these figures that a full scrubbing option (e.g., the proposed 1.2 ceiling, 0.2 floor, 85 percent removal, 24-hour average) will promote the greatest use of local coal and minimize the shipment of western coal to the East.

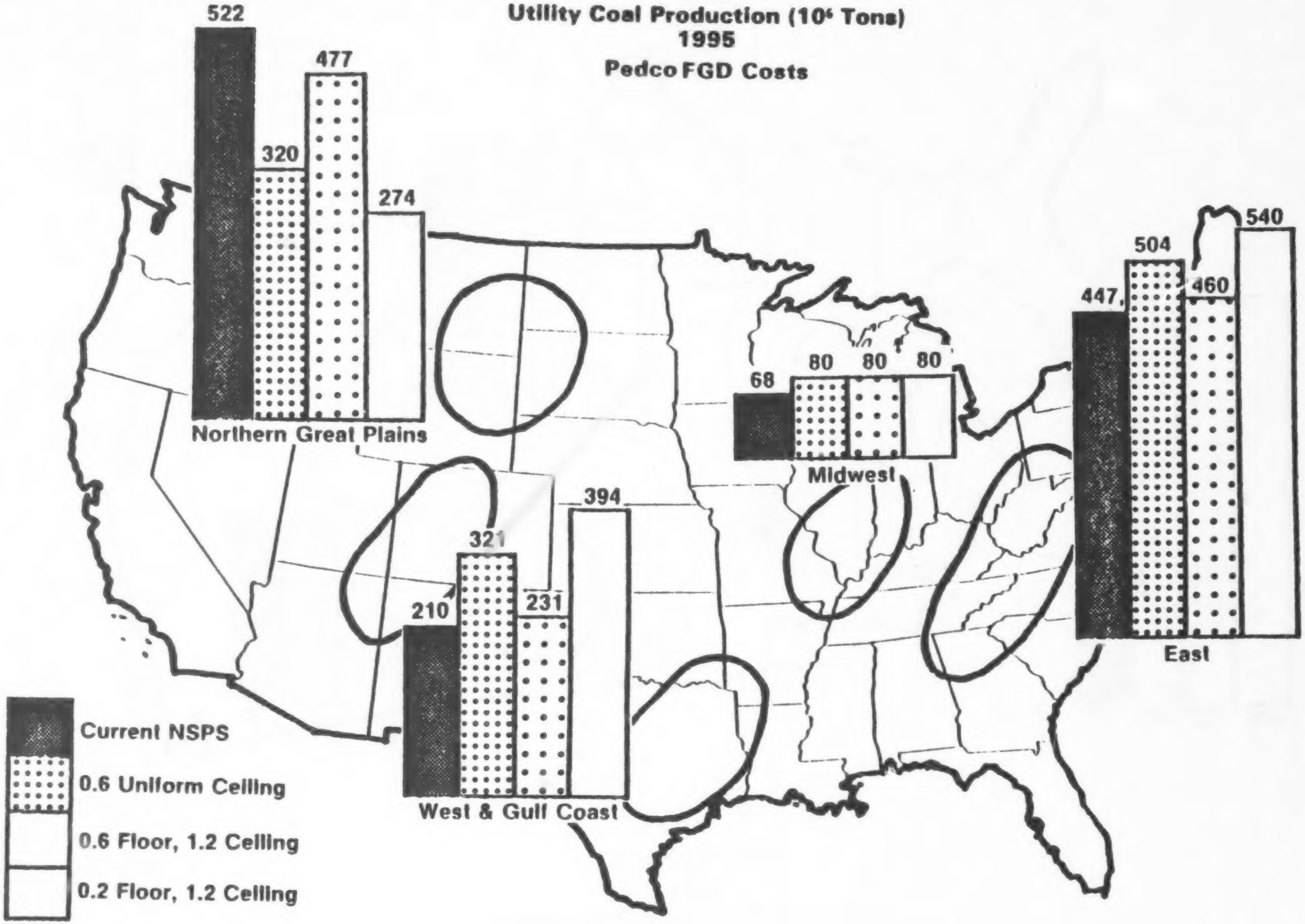
Regional FGD capacity in 1995 is shown in Figure 15 for the high fuel cycle cost case. In the low fuel cycle cost case FGD capacity will be higher under the current NSPS and under partial scrubbing options, particularly in the central "swing" states which are more sensitive to changes in relative fuel-cycle costs.

Estimated FGD sludge, coal ash, FGD capacity, and utility water consumption in 1995 are shown in Figures 16 and 17 for the high fuel cycle cost case. It is clear from these figures that sludge production and FGD water consumption are relatively insensitive to the form of the revised NSPS. In absolute terms FGD sludge quantities will be of the same order of magnitude as fly ash quantities and FGD water consumption is projected to be an order of magnitude less than consumptive cooling water requirements. Of course, the environmental impacts of sludge and water consumption will depend on specific power plant locations.

Utility fossil fuel consumption in 1995 for the high fuel-cycle cost case is shown in Figure 18. Note that projected oil consumption is independent of the revised NSPS and that fuel used in coal transportation is an order of magnitude less than boiler oil consumption. Oil consumption is independent of the SO_2 standard because of our plant retirement and dispatching strategy. Considerable oil plant retirements are projected to occur in the decade between 1985 and 1995. However, in our model, oil plants are retired on the basis of age, announced utility plans, government coal conversion programs, and not strictly economics. We feel that this is appropriate for a number of reasons including:

- High fuel oil costs are usually passed through to the customer.

Figure 11
Utility Coal Production (10^6 Tons)
1995
Pedco FGD Costs



522
320
477
274
Northern Great Plains
68
80
80
80
Midwest
447
504
460
540
East
210
321
231
394
West & Gulf Coast
Current NSPS
0.6 Uniform Ceiling
0.6 Floor, 1.2 Ceiling
0.2 Floor, 1.2 Ceiling

Figure 12
Utility Coal Production (10^6 Tons)
1995

TVA FGD Costs

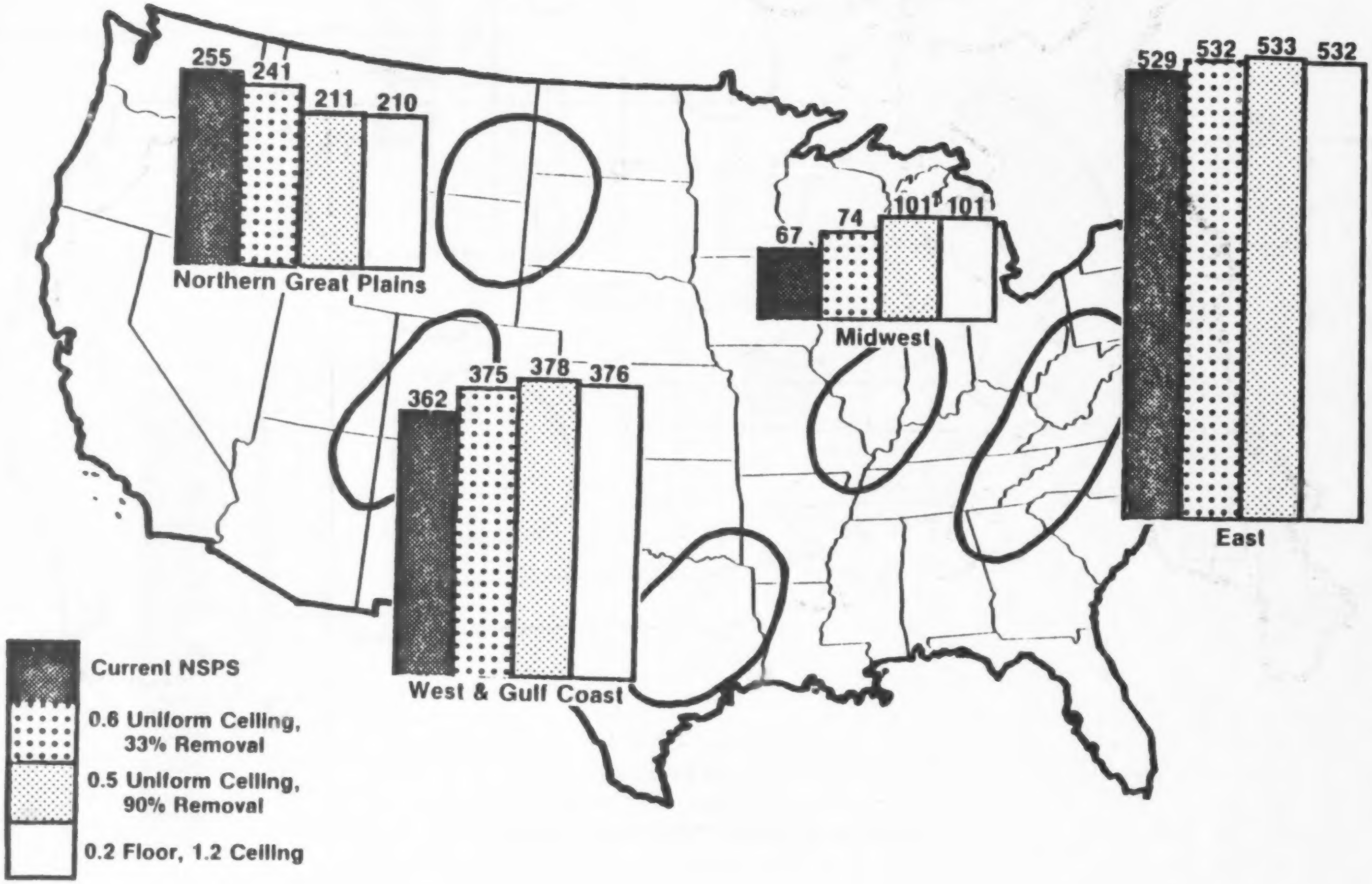

255
241
211
210
Northern Great Plains
67
74
101
101
Midwest
362
375
378
376
West & Gulf Coast
529
532
533
532
East
Current NSPS
0.6 Uniform Ceiling, 33% Removal
0.5 Uniform Ceiling, 90% Removal
0.2 Floor, 1.2 Ceiling

Figure 13
Western Coal Shipped East
1995
Pedco FGD Costs

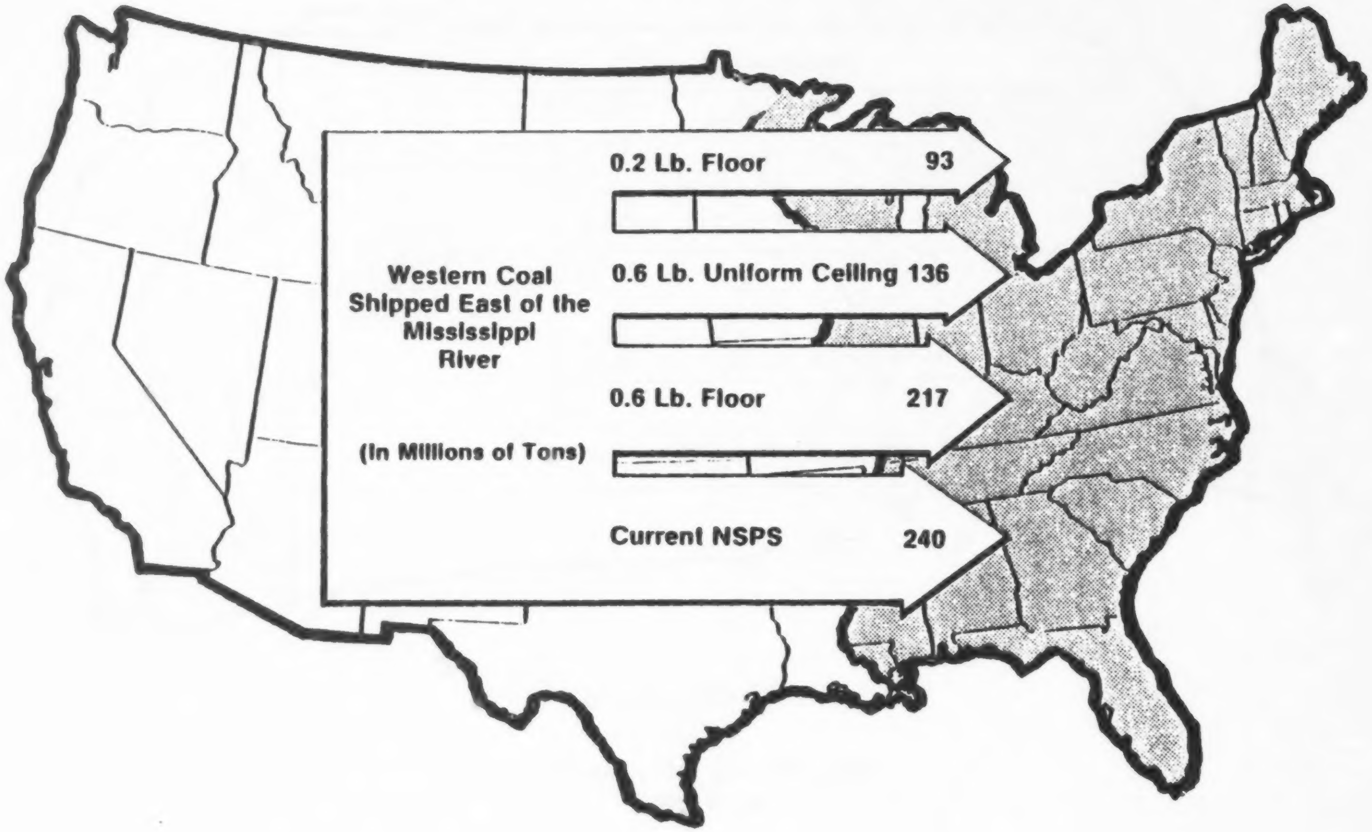
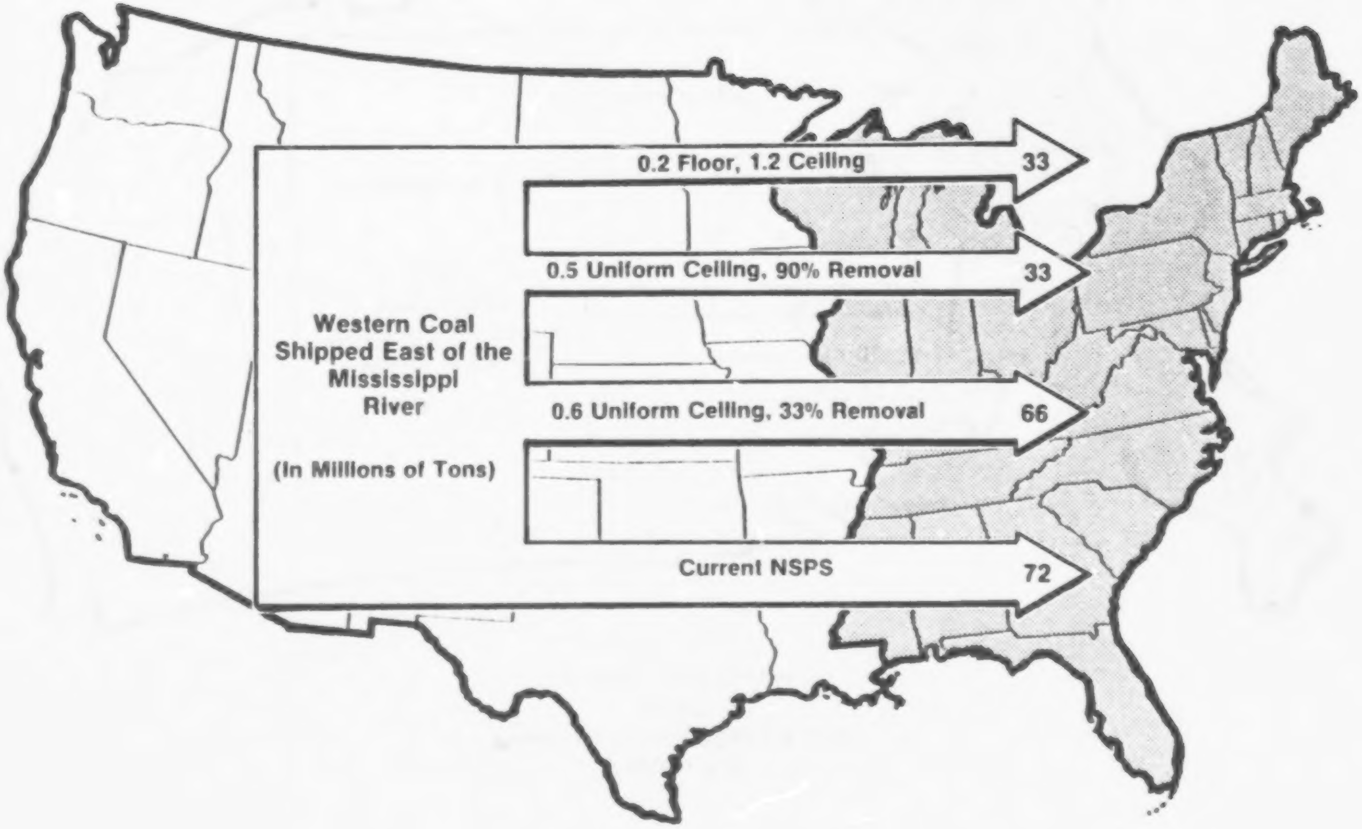

Western Coal Shipped East of the Mississippi River
(In Millions of Tons)
0.2 Lb. Floor 93
0.6 Lb. Uniform Ceiling 136
0.6 Lb. Floor 217
Current NSPS 240

Figure 14
Western Coal Shipped East
1995

TVA FGD Costs



0.2 Floor, 1.2 Ceiling
33
0.5 Uniform Ceiling, 90% Removal
33
Western Coal Shipped East of the Mississippi River
0.6 Uniform Ceiling, 33% Removal
66
(In Millions of Tons)
Current NSPS
72

Figure 15
Regional FGD Capacity (GW)
1995
Pedco FGD Costs

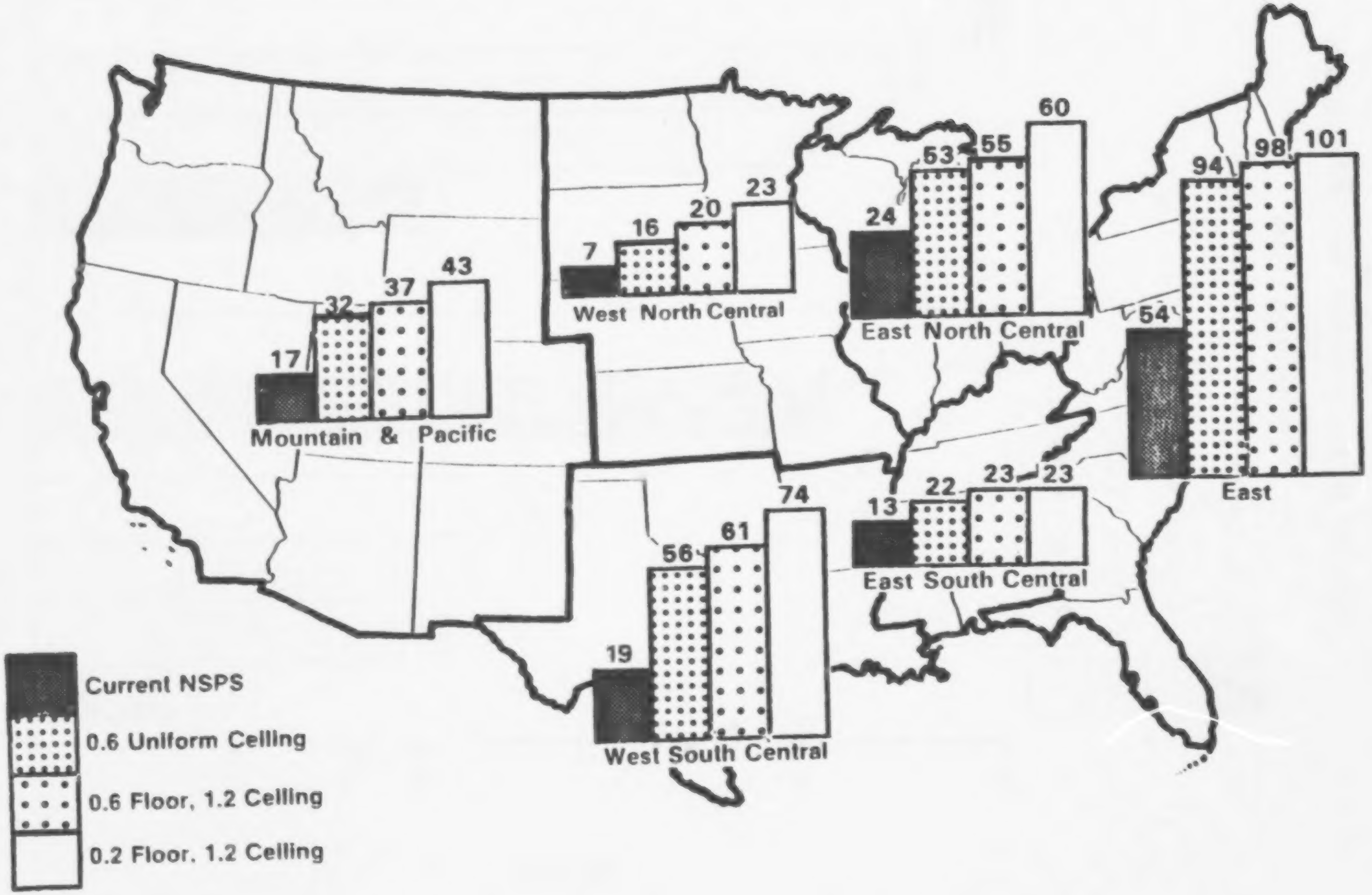

17
32
37
43
Mountain & Pacific
7
16
20
23
West North Central
24
53
55
60
East North Central
54
94
98
101
East
13
22
23
23
East South Central
19
56
61
74
West South Central
Current NSPS
0.6 Uniform Ceiling
0.6 Floor, 1.2 Ceiling
0.2 Floor, 1.2 Ceiling

Figure 16
National Sludge, Coal Ash, and FGD Capacity
1995
Pedco FGD Costs

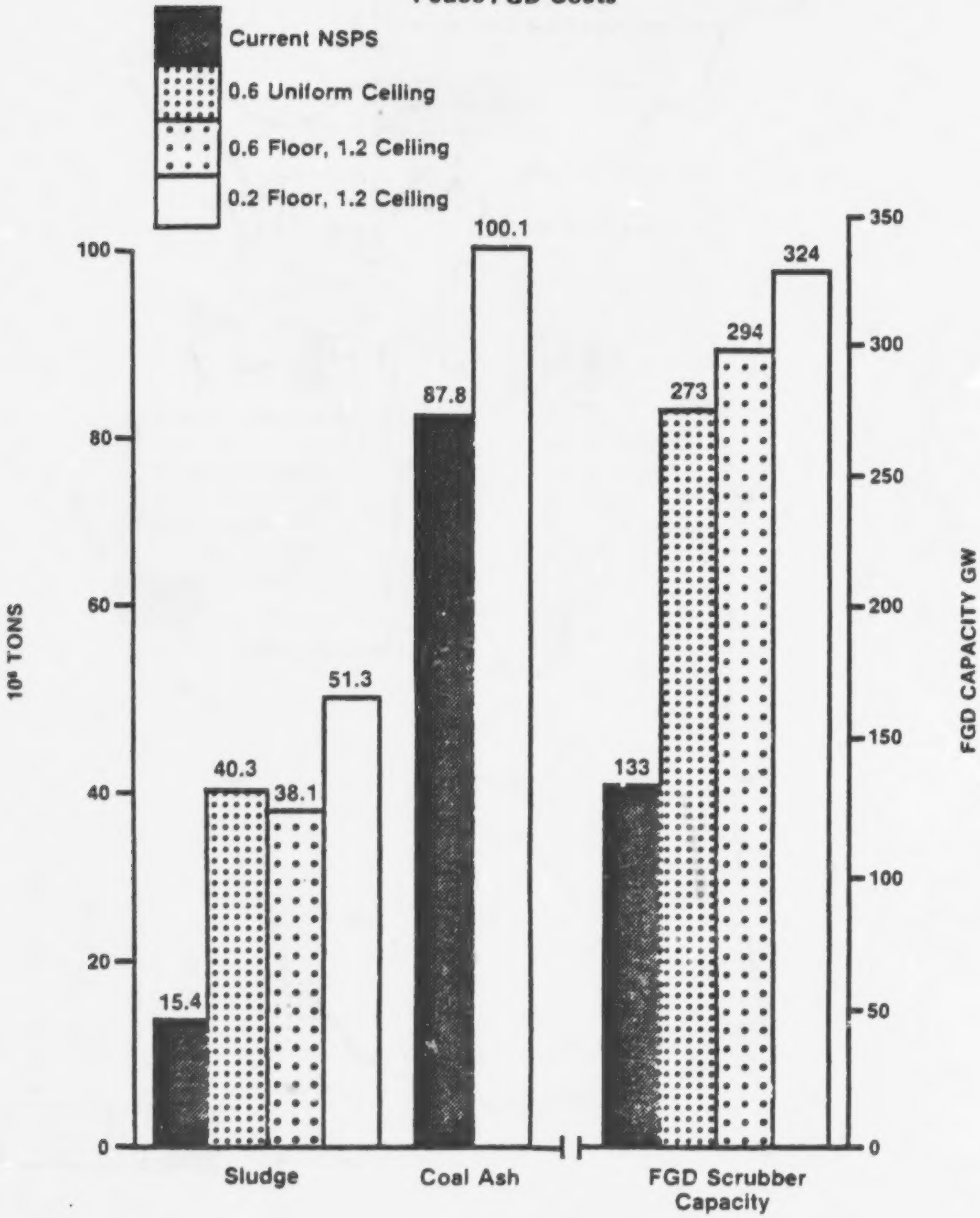


Figure 17
Utility Water Consumption
1995

Pedco FGD Costs

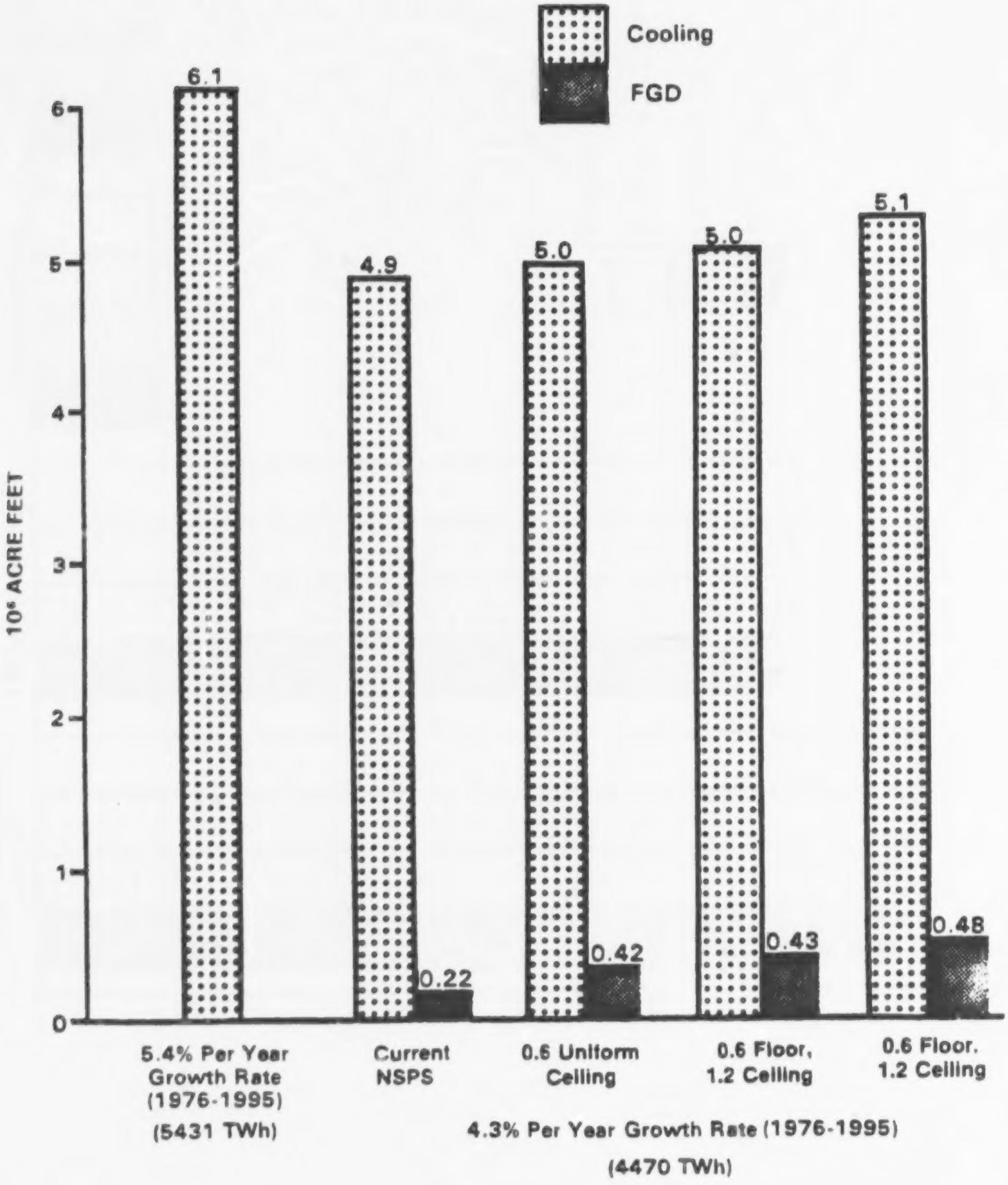
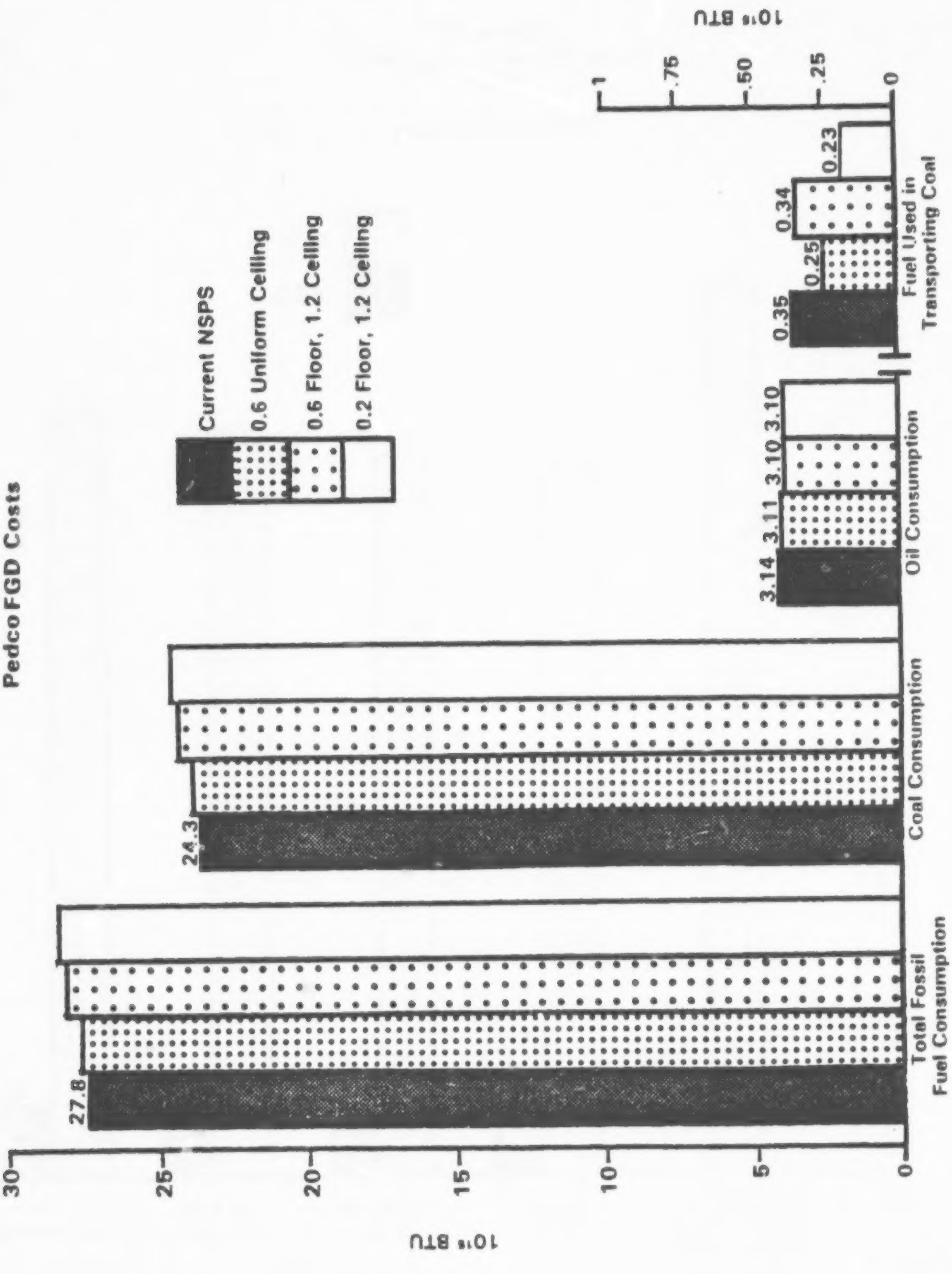


Figure 18
Utility Fossil Fuel Consumption
1995
Pedco FGD Costs



- Oil plants are often located in urban areas where coal storage space is not available.
- It is much easier for a utility to operate an existing oil plant than to site, build, and operate a new coal plant.
- Oil plants are often located in strategic locations in the distribution grid and in 1995 will be used in a cycling mode.
- Residual oil will be available as long as petroleum is refined for gasoline for use in motor vehicles, etc.
- Lower reserve margins in 1995 (about 20 percent) will discourage differential retirements of the remaining oil capacity simply in response to more stringent RNSPS.

Oil plants in the 1990's will be dispatched after coal plants because of their high operating cost. Since the demand curves are assumed constant for each alternative NSPS, their use, and hence oil consumption, does not change with the alternative New Source Performance Standards. If, however, scrubber reliability is lower than assumed, oil plants could be utilized more – although it is also possible that utilities might build more nuclear plants if this were the case.

The costs associated with alternative New Source Performance Standards are illustrated in Figures 19–21. The pollution control investment in Figures 19 and 21 are for all pollution control investments including those for particulate control, water pollution control, and SO_2 control. Figures 19 and 20 show that the uncertainties in projected fuel-cycle cost (i.e., PEDCo vs TVA FGD costs) can lead to substantial differences in estimated costs. Total pollution control investment between 1983 and 2000 is compared to total utility investment in Figure 21 and illustrates that pollution control investment will represent less than ten percent of the total utility investment during that period.

Cost Effectiveness of a Standard

The cost effectiveness of alternative revised New Source Performance Standards can be measured in numerous ways, some of which are not especially

Figure 19

Comparison of Cumulative Pollution Control Investment 1983-2000 Reflecting Pedco and TVA FGD Costs (Billions 1975$)

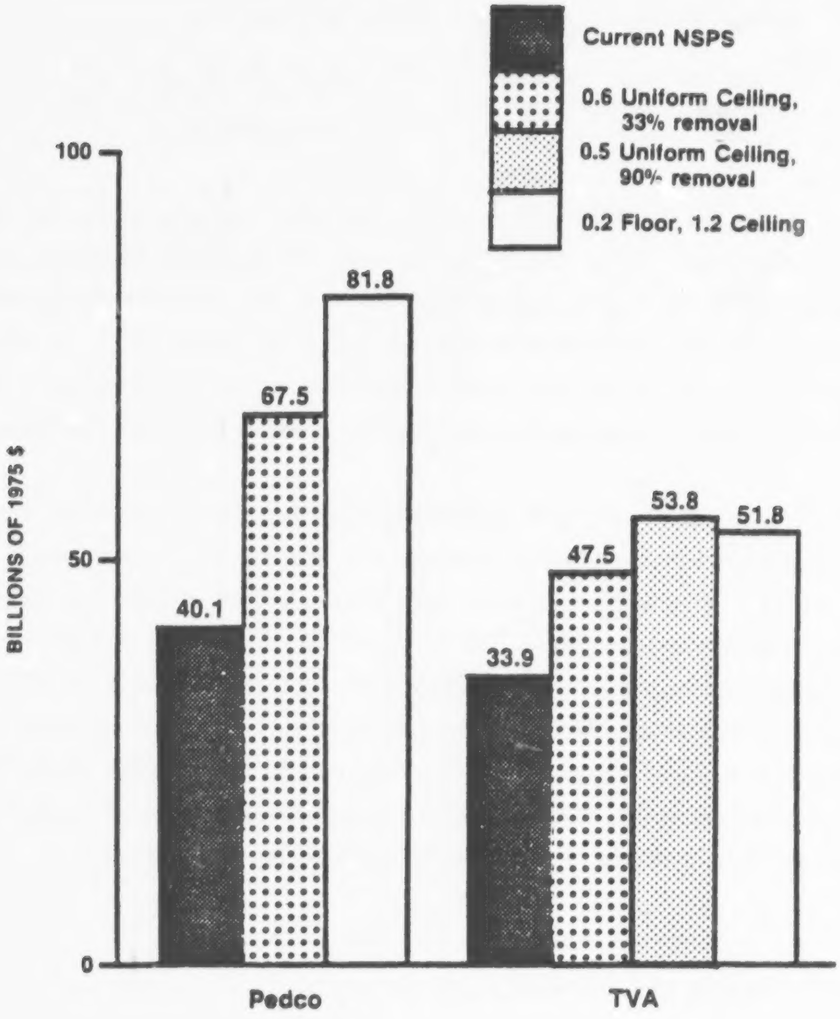


Figure 20

National Average Residential Monthly Electric Bill in 1995 and Percentage Increase from Current NSPS (1975$)

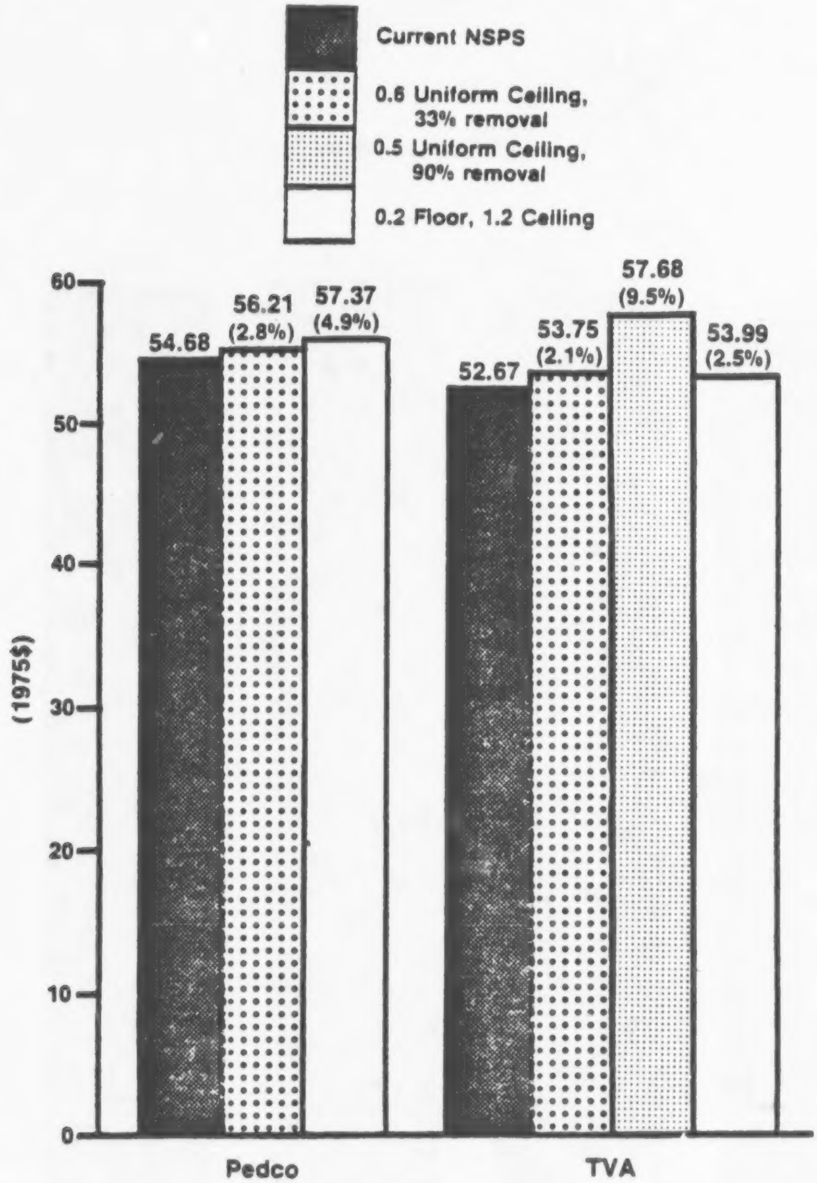
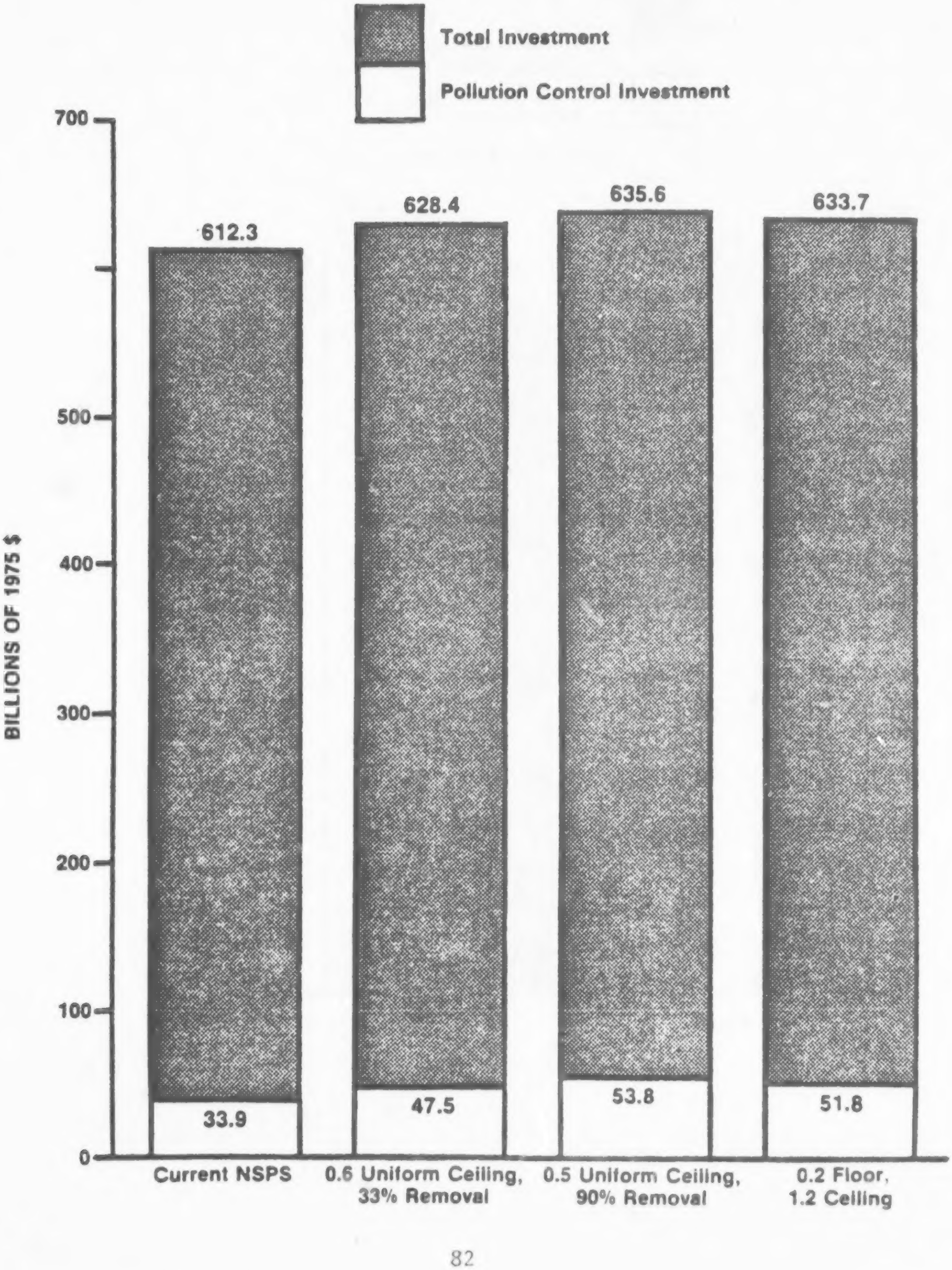


Figure 21
Comparison of National Pollution Control Investment and Total Cumulative Investment 1983-2000 (Billions 1975 $)

TVA FGD Costs



useful in distinguishing between alternative standards. The different interpretations of cost effectiveness measures are illustrated in Figures 22 and 23. If FGD costs in cents per million Btu are examined, as in Figure 22, it is clear that it is less expensive to remove SO_2 from low sulfur coals to achieve a given emission limit. If this cost effectiveness measure is used (which relates directly to the cost per kilowatt-hour for electricity), then a standard favoring the use of low sulfur coals should be established. On the other hand, if cost effectiveness is measured in terms of dollars per ton of SO_2 removed, as in Figure 23, it is clear that the use of high sulfur coal provides the greatest cost effectiveness. Therefore, the most cost effective standard by this measure would be one that promotes the use of high sulfur coal. Therein lies the dilemma presented to decision makers who must ultimately select the revised New Source Performance Standard. Also therein lies the challenge to the engineers and technologists gathered at this conference who are working to improve each of these cost-effectiveness measures by reducing the future costs of SO_2 removal.

Figure 22
Comparison of FGD Cost Effectiveness per Btu of Fuel Input Under Annual Average SO_2 Control Alternatives

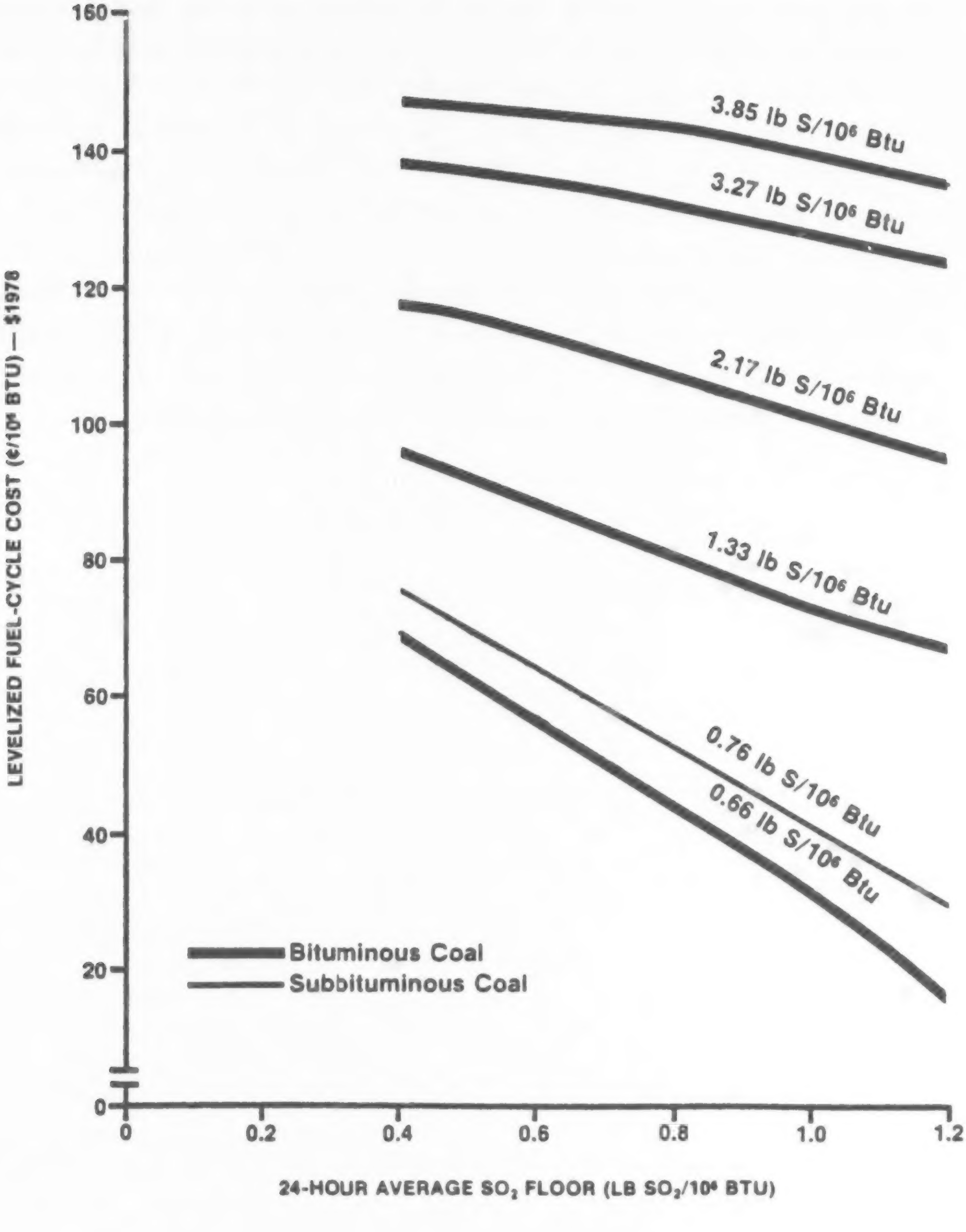
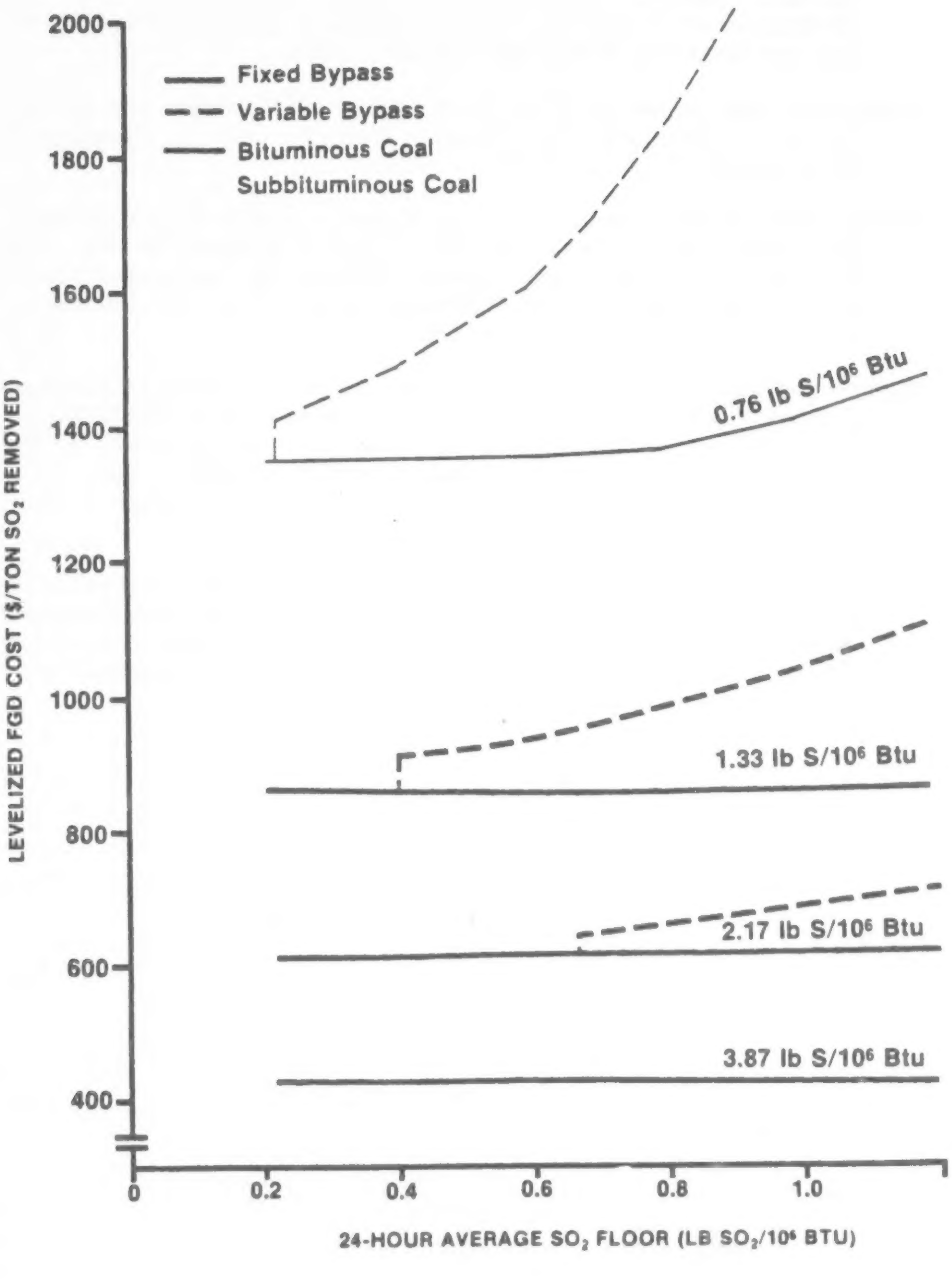


Figure 23
Comparison of FGD Cost Effectiveness per Ton of SO2 Removed under 24-Hour Average SO2 Control Alternatives with a 1.2 lb/10₆ Btu Ceiling



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2. "Additional Information on EPA's Proposed Revision to New Source Performance Standard for Power Plants," Federal Register 43 (8 December 1978): 57834-59.

3. PEDCo Environmental, Inc., "Summary Report – Utility Flue Gas Desulfurization Systems, Oct.-Nov. 1977," Report prepared for the U.S. Environmental Protection Agency, Division of Stationary Source Enforcement and Industrial Environmental Research Laboratory (Cincinnati, Ohio, 25 January 1978).

4. PEDCo Environmental, Inc., "Particulate and Sulfur Dioxide Emission Control Costs for Large Coal-Fired Boilers," EPA-450/3-78-007, Prepared for U.S. Environmental Protection Agency, Office of Air Quality Planning and Standards, Research Triangle Park, N.C. (Cincinnati, Ohio, February 1978). Includes detailed computer printouts for all case studies.

5. TVA-Bechtel Shawnee Limestone-Lime Computer Program: ten printouts (lime 25 MW, 100 MW, 200 MW, 500 MW, 1000 MW; and limestone 25 MW, 100 MW, 200 MW, 500 MW, 1000 MW). Provided by C. David Stephenson, National Fertilizer Development Center, Muscle Shoals, Alabama, December 1978.

SESSION 2

IMPACT OF RECENT LEGISLATION

WALTER C. BARBER, CHAIRMAN

Panel: Impact of Recent Legislation
Brief overviews of recent legislation, EPA's approach to implementation, and potential impacts followed by questions from the audience.

Members: James L. Agee
Gary N. Dietrich

No papers or discussions are included for this session.

SESSION 3

ECONOMICS AND OPTIONS

WALTER C. BARBER, CHAIRMAN

PAPER 3A

STATUS OF DEVELOPMENT, ENERGY AND ECONOMIC ASPECTS OF ALTERNATIVE TECHNOLOGIES

P. S. Farber, C. D. Livengood,
K. E. Wilzbach, W. L. Buck, and H. Huang
Argonne National Laboratory

Several energy technologies are under development throughout the world that either totally negate the need for flue-gas desulfurization (FGD) or require less than full flue-gas scrubbing. These processes remove sulfur either prior to coal combustion (coal cleaning or conversion), during combustion (atmospheric and pressurized fluidized-bed combustion), or "between" two combustion stages (gasification/combined-cycle operation).

This paper reviews the status of development and/or demonstration of these technologies with respect to their possible application to the generations of electricity. In addition, the overall coal-to-electrical-energy conversion efficiency and economics (capital costs and total annualized costs, mills/kWh) are explored and compared for the various alternatives. The economic premises utilized conform, as much as possible, to those used by the TVA in comparisons of FGD technology. The paper shows, among other things, the importance in any energy and economic analysis of an energy system, be it a postcombustion treatment process (FGD) or the total energy process (FBC), of taking into account the cost of fuel and the overall process energy efficiency.

1 INTRODUCTION

Over the last decade there has been increasing interest in developing alternatives for the coal-to-electricity process other than conventional combustion. The impetus for these efforts has come from two sources, the OPEC oil embargo of the early seventies, which led to increased emphasis on coal use, and environmental regulations such as the current New Source Performance Standards for utility power plants. As a consequence, the process development efforts have emphasized increased overall efficiency and burning of coal in an environmentally acceptable manner. Recently, the proposed tightening of NSPS restrictions on sulfur-oxide emissions, coupled with doubts as to the ability of conventional flue-gas desulfurization to achieve a high degree of reliability together with an 85-90% removal efficiency, has brought an even greater awareness in government and utility circles of the need to develop power plant cycles of greater efficiency and minimum environmental impact. To be considered as commercially acceptable, however, these power plants must produce electricity at costs competitive with, or less than, conventional combustion with flue-gas desulfurization (FGD).

Three differing energy technologies that will either negate the need for flue-gas desulfurization or require less than full flue-gas scrubbing are under testing and development by industry and the government. These processes may be broken down into three categories: (1) those that remove sulfur prior to combustion (coal conversion), (2) those where the sulfur removal takes place during combustion (atmospheric and pressurized fluidized-bed combustion), and (3) those where sulfur removal takes place between two combustion stages (gasification/combined-cycle power systems).

As part of a program sponsored by the Department of Energy, Argonne National Laboratory has been evaluating the environmental and economic aspects of these alternative technologies with respect to their application to electricity generation. This paper reviews the status and background of these technologies and analyzes their relative economic attractiveness. In order to allow a direct comparison with conventional coal to electricity systems (with FGD), the economic premises used conform, as much as possible, to those used by the TVA. A comparison by TVA of various processes used in flue-gas desulfurization follows this paper in the symposium.

2 TECHNOLOGY CHARACTERISTICS

2.1 SOLVENT-REFINED COAL

The solvent-refined-coal (SRC) process has received considerable attention during the past several years, largely because end products from this process can possibly be used in fuel-combustion sources to comply with environmental standards. At present, two pilot plants are in operation: A 6-ton/day unit (since 1973) by Southern Company Services at Wilsonville, Alabama, and a 50-ton/day unit (since 1974) by the Pittsburgh and Midway Coal Mining Company of Gulf Oil at Fort Lewis, Washington. The former has been sponsored by EPRI and DOE, with the primary objective being collection of operating information on the main reactor (dissolver) and several solid-liquid separation devices, using a number of different coals. On the other hand, the operation of the larger plant, sponsored principally by DOE, has emphasized collecting technical data on a Kentucky coal to validate scale-up to commercial production, and to provide large samples of SRC for combustion and market-development studies.

2.1.1 Process Description

Basically, two operating modes of the SRC process have been identified.[1] The original mode (known as SRC-I) is intended, with minimum hydrogen consumption, to produce low-sulfur, low-ash solid products that are suitable for use as boiler fuel without flue-gas-treatment provisions. The slurry-recycle mode, considered as an advanced version and known as SRC-II, consumes more hydrogen and produces a low-sulfur, ash-free fuel oil, along with significant amounts of light oil, pipeline gas, and naphtha. Although conclusive economic figures are not available, it is generally believed that the solid SRC should be less costly (per Btu) than its liquid counterpart.

Figure 1 shows a simplified schematic of the SRC-I process. The coal is first pulverized and mixed with a coal-derived, anthracene-oil-type solvent in a slurry-mixing tank. The slurry is mixed with hydrogen (produced elsewhere in the process) and is then pumped through a fired preheater and into a dissolver, where about 85-95% of the coal (moisture and ash-free) is dissolved. The process operating conditions vary with the type of coal that is processed; some representative values are given in Table 1. Under these conditions, several reactions -- depolymerization and hydrodesulfurization -- also occur, resulting in a very complex solid-liquid-gas mixture.

From the dissolver, the mixture flows to a separator, where the gases are separated from the slurry of undissolved solids and coal solution. The raw gas is directed to a desulfurization unit, then to a hydrogen-recovery unit. The hydrogen sulfide is converted to elemental sulfur, hydrocarbon gases are given off, and the recovered hydrogen is recycled. The slurry of undissolved solids and the coal solution flows to a solid-liquid separator, where the solids are removed. The solid residue, composed mainly of unconverted carbon and ash, is sent along with supplemental coal to a gasifier/converter to produce make-up hydrogen. The coal solution flows to a vacuum

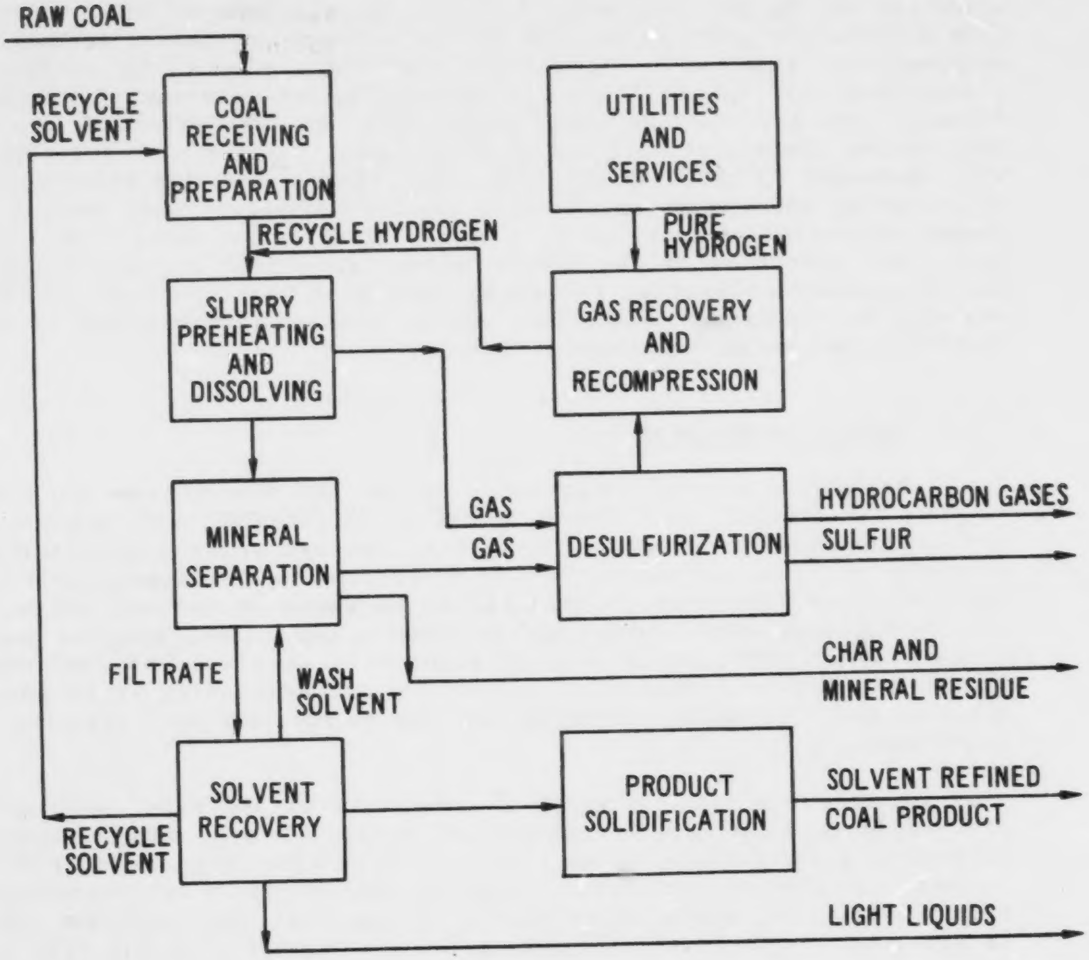

RAW COAL
RECYCLE SOLVENT
COAL RECEIVING AND PREPARATION
UTILITIES AND SERVICES
RECYCLE HYDROGEN
PURE HYDROGEN
SLURRY PREHEATING AND DISSOLVING
GAS RECOVERY AND RECOMPRESSION
GAS
GAS
MINERAL SEPARATION
DESULFURIZATION
HYDROCARBON GASES
SULFUR
CHAR AND MINERAL RESIDUE
FILTRATE
WASH SOLVENT
SOLVENT RECOVERY
PRODUCT SOLIDIFICATION
SOLVENT REFINED COAL PRODUCT
RECYCLE SOLVENT
LIGHT LIQUIDS

Figure 1. SRC-I Process Schematic

Table 1. SRC Operations at Ft. Lewis

Process Parameter	Typical Value
Coal Source:	
State	Kentucky
Mine	Colonial
Seam	#9 and #14
Sulfur (%)	3.1
Temperature (F)	850
Pressure (psig)	1500
Coal Conversion (% MAF Coal)	95
SRC Output:	
Yield (% MF Coal)	65
Sulfur (%)	0.8
Gas Yield (% MF Coal)	6.0
Liquids (% MF Coal)	13.5
Solvent (% MF Coal)	4.4

still for recovery of light oils and recycle solvent. The SRC is produced by solidification. The SRC has a melting point of 350 to 450 F and a heating value of about 16,000 Btu/lb.

2.1.2 Product Characterization

Large quantities of blended Kentucky coals have been processed at Fort Lewis in both the original SRC and slurry-recycle modes. This coal is a blend of #9 and #14 seams in Hopkins City, Kentucky, as obtained from the Colonial #1 mine of P&MCM Co. Available data regarding the physical/chemical properties of solid SRC samples as well as the raw coal have been collected in a report[2] prepared for ANL by Air Products and Chemicals and are presented in Table 2. Also shown in Table 2 for comparison are typical properties of liquid SRC oil obtained with a similar coal feedstock.

It is obvious that SRC is a better boiler fuel than the raw coal in terms of ash, sulfur, and heat content. Also, storage and handling tests by Air Products[3] and Babcock and Wilcox[4] did not uncover any insurmountable problems. Nevertheless, it appears that the use of solid SRC in utility boilers without additional controls to meet the proposed NSPS standards (85% sulfur removal and 0.03 lb TSP/10^6 Btu) is questionable, though the current standard for SO_2 can be met. (Utility-scale combustion tests are discussed in the following section.) Further process improvement or product upgrading is imperative to maintain SRC as a compliance boiler fuel in the future.

Table 2. Properties of Ft. Lewis SRC and Raw Coal

Property	SRC Solid	Coal	SRC Oil[a]
A. Proximate Analysis (wt % dry)			
Volatile Matter	61.4-66.7	39.8	
Fixed Carbon	33.3-38.5	49.8	
Ash	0.08-0.12	10.4	
B. Ultimate Analysis (wt % dry)			
Carbon	87.1-87.7	70.4	86.6
Hydrogen	5.5-5.9	5.1	8.4
Nitrogen	1.8-2.0	1.4	1.1
Sulfur	0.7-0.9	3.4	0.3
Ash	0.08-0.12	10.4	0.01
Oxygen (by difference)	3.98-4.22	9.2	3.6
C. Heating Value (Btu/lb dry)	15770-15810	12,760	17,040
D. Sulfur Forms (wt % dry)			
Pyritic	0.014-0.019	1.5	
Sulfate	0.002-0.004	0.2	
Organic	0.684-0.877	1.7	
Total	0.700-0.900	3.4	
E. Hardgrove Grindability Index (HGI)		55-70	
Temperature (F)	HGI		
10	170-178		
70	184-188		
150	180-187		

[a] API Gravity (60 F): 8.3, and viscosity (SSU at 140 F): 35.6

These are some of the main reasons for the development of the slurry-recycle SRC process. The liquid SRC oil is claimed to be similar to No. 4 fuel oil in terms of handling and burning and can be used to comply with the proposed NSPS, albeit at higher costs.

2.1.3 Combustion Tests

About 3,000 tons of solid SRC were burned in a 22.5-MWe boiler at Georgia Power Company's Mitchell Plant in mid-1977[5], and about 4,500 barrels of liquid SRC oil were consumed at the Consolidated Edison's 74th-Street station in Manhattan in late 1978[6]. The Con. Ed. test was reported to be a success, and all air emissions were below the newly proposed NSPS for utility boilers using coal-derived fuels. However, the fuel characteristics were not the same as those expected for commercial operation, having a higher ratio of light to heavy oils. Bench-scale burn tests, nevertheless, have demonstrated that the proposed NSPS for boilers using coal-derived fuels could be met[7] with the use of staged combustion to suppress NO_x.

No serious problems were encountered in the combustion test of solid SRC at the Mitchell plant.[5] Blowing losses experienced in early rail-car shipments and dusting during unloading operations were successfully minimized by simple chemical pretreatments. The standard pulverizers were modified to the extent of using unheated air, reducing ball-spring pressure, and installing variable-speed feeder motors. The only major boiler modification required was the use of specially designed, water-cooled dual-register burners to accomodate the low melting point of solid SRC and to reduce NO_x emission.

Operation and emissions data are reported in Reference 5 for solid SRC and raw coal. No CO, C_1-C_6 hydrocarbons, or polynuclear aromatic compounds were detected during any SRC test, and overall boiler efficiency at full load was essentially the same when burning either SRC or coal. Of the three most important air pollutants, it appears that only NO_x emissions were below both current and proposed NSPS (0.5 lb $NO_2/10^6$ Btu) for utility boilers (by about 40% and 10%, respectively). SO_2 emissions were about 20% below the current NSPS but barely fall short of the 85% removal mark. The particulate loadings leaving the boiler were surprisingly high due to the high carbon content of the fly ash and the unexpected higher-than-normal ash content in SRC (0.57 wt percent) resulting from contamination by surface dust and other foreign material during pit solidification and storage. Even with a secondary precipitator of modern design, the plant could not meet the newly proposed NSPS for particulates, probably again due to the high carbon concentration of SRC fly ash. Also note that the 20% opacity standard was not met under full and medium loads.

In mid-1978, DOE awarded two $6-million contracts to consortiums headed by Southern Company Services and Gulf Oil to start the preliminary designs of two 6,000-ton-per-day plants to make solid SRC and liquid SRC oils, respectively. Both design studies are scheduled to be finished about mid-1979. At that time, a choice is scheduled to be made as to which type of plant will actually be built.

2.2 GASIFICATION/COMBINED-CYCLE POWER GENERATION

The combination of coal gasification with combined-cycle power generation, utilizing both a gas turbine and waste-heat boiler, offers the promise of economically competitive power generation together with highly effective pollution control. However, these possible advantages are obtained at the expense of system simplicity, and many unanswered questions regarding the operability and economics of full-scale systems remain. The results presented in this study were obtained from an assessment of the technology prepared for ANL by the United Technologies Research Center.[8]

2.2.1 Process Description

The gasification/combined-cycle power plant is an integrated facility: It includes all the equipment necessary to convert coal into electricity. As can be seen in Figure 2, the conversion process can be divided into two major systems, fuel processing and power production.

The fuel-processing system includes coal storage and handling, coal processing, gasification, fuel treatment, and by-product streams. Coal is delivered from the coal yard to a crusher, where it is sized to dimensions required by the gasifier, e.g., 1/8 x 1-1/2 in. for a Lurgi to pulverized coal (70% through 200 mesh) for a Foster Wheeler gasifier. From the crusher, the coal can be dried, if necessary, and then injected into the gasifier.

The resultant fuel gas, having a heating value from 90 to 150 Btu/scf, is now sent to the fuel-gas treatment portion of the procesing plant. Here the gas is cooled, and tars, if present, are separated along with particulates and ammonia and sent to the sulfur-removal process, where the H_2S, COS, and other sulfur compounds are removed. The clean fuel gas continues on to the power-production portion of the power plant while the sulfur bearing gases are sent to a Claus plant for conversion to elemental sulfur.

The major by-products of the fuel-processing system are elemental sulfur, ammonia, slag and/or ash, and, for some gasifiers, coal tars.

The power-production section consists of gas turbines, heat-recovery steam generators, steam turbogenerators, and heat-rejection equipment. The fuel gas from the fuel-processing section is burned in the gas turbine. A portion of the compressor discharge of the gas turbine is sent to the gasification plant via a boost compressor to supply oxidant for the process. The hot exhaust gases from the gas turbines are cooled by raising steam, which is sent to a steam turbine, expanded to subambient conditions producing power, and condensed. The heat of condensation is rejected in mechanical draft cooling towers. Steam may also be produced by cooling the fuel gas prior to sulfur removal.

There are a number of auxiliary systems required in both the fuel-processing and power-production facilities. In addition to the normal stationkeeping requirements for heat, light, potable water, compressed air, and

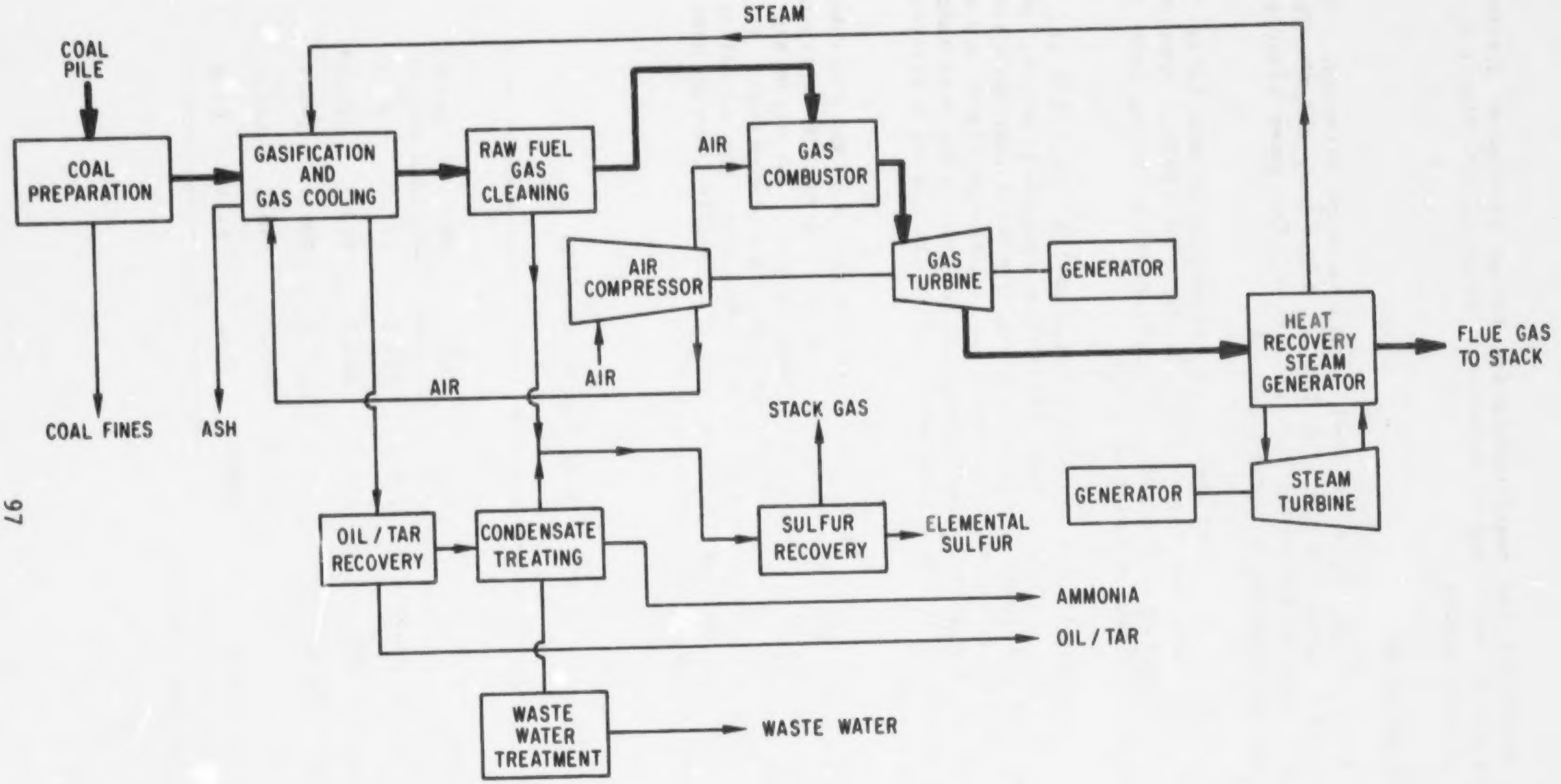

STEAM
COAL PILE
COAL PREPARATION
GASIFICATION AND GAS COOLING
RAW FUEL GAS CLEANING
AIR
GAS COMBUSTOR
AIR COMPRESSOR
GAS TURBINE
GENERATOR
HEAT RECOVERY STEAM GENERATOR
FLUE GAS TO STACK
COAL FINES
ASH
AIR
AIR
STACK GAS
GENERATOR
STEAM TURBINE
OIL / TAR RECOVERY
CONDENSATE TREATING
SULFUR RECOVERY
ELEMENTAL SULFUR
AMMONIA
OIL / TAR
WASTE WATER TREATMENT
WASTE WATER

Figure 2. Gasification/Combined-Cycle Power-Plant Schematic

instrumentation, there are the requirements for boiler feed-water treatment, cooling-water treatment, waste-water treatment, disposal of slag/ash, and transfer/storage of by-products.

2.2.2 Gasifier Descriptions

For this study, four types of coal gasifiers were selected. These include the pressurized Lurgi dry-ash, the IGT-Gas fluid-bed, and the Texaco and two-stage Foster Wheeler entrained-flow gasifiers. Key power-plant parameters are given for each gasifier type in Table 3.

The Lurgi dry-ash gasifier has been commercially available from Lurgi Gesellschaft fur Warme und Chemie Technik MbH since 1936. Presently, there are nearly 70 gasifiers in commercial operation producing town gas, synthesis gas (syn gas) or medium-Btu fuel gas.

In this system, coal is crushed to between 1/8 and 1-1/2 in. and sent to one of two lock hoppers. Coal fines are usually rejected, which results in as much as a 25% loss in delivered coal. However, consideration is being given to briquetting fines with recovered tar. Although claims are made that all coals, including highly caking types, can be used, some pretreatment would probably be required for the caking coals. Also the use of a stirrer in the bed could be necessary.

Coal from the lock hopper is dropped into the gasifier and distributed evenly by the distributor arm on a coal bed. Steam and air are introduced into the bottom of the gasifier through a rotating grate. The steam and air pass upward through the bed, creating different zones in the gasifier. At the bottom, carbon is burned, providing heat to the next zone, the gasification zone. The hot gases then devolatilize the coal and finally provide heat to dry the incoming coal prior to leaving the gasifier.

Table 3. Power System Characteristics

Power System	Lurgi	IGT	FW	Texaco
Clean Gas HHV (Btu/scf)	109.4	158.6	177.9	90.5
Gas Turbine Pwr (MW)	450.9	328.2	320.7	273.0[a]
Steam Turbine Pwr (MW)	122.1	237.6	206.4	194.4
Net Pwr (MW)	532.8	523.2	488.1	441.2
Heat Rate (Btu/kWh)	10869	8264	8304	8706

[a] Includes let-down turbines.

Considerable amounts of tar and oils are produced in the upper zones and must be removed from the gas. This is accomplished in a scrubber cooler, a waste-heat boiler and subsequent coolers. The separated tar liquors and oils are usually fed to a Phenosolvan unit and ammonia plant to recover crude phenols and anhydrous ammonia.

The Lurgi gasifier operates at a pressure that is a function of gas-turbine pressure ratio and fuel-processing-system and fuel-control-system pressure drop. The commercially available gasifier can process approximately 400 ton/day and has an expected turndown ratio of 4 to 1.

The U-Gas process is being developed by the Institute of Gas Technology, where a 4-ft-diameter atmospheric-pressure unit processing coke has been in operation since 1974. This process utilizes a fluidized-bed system that can produce either low- or medium-Btu gas with either air- or oxygen-blown operation.

The use of a fluidized bed has many inherent advantages. In particular, the bed material acts as a catalyst for the gasification reactions and should permit operation at relatively low temperature while completely gasifying the feed. The U-gas process is distinguished from other fluid-bed processes in that it utilizes an "ash agglomeration" technique to concentrate the ash and remove it with minimum carbon content while operating the bed with a relatively high carbon content.

Raw coal is crushed to 0 x 1/4 in. size. The feed may contain up to 10% <200-mesh material as generated in the crushing step. Noncaking, subbituminous coals and lignite can be fed directly to the gasifier from the crusher. Caking coals (eastern bituminous, for example) must at present be pretreated by contact with air in a fluidized bed operating at gasifier pressure and 700 to 800 F. An oxidized outer layer forms on the coal particles, preventing agglomeration and possible blockage in the gasifier.

Heat evolved during pretreatment is removed by generating steam in heat-transfer coils immersed in the fluidized-bed pretreater. Coal that has been pretreated is fed to the gasifier. Off-gases are fed to the bottom of the gasifier to destroy all tar and oils that evolve during the pretreating process.

The gasifier is a refractory-lined, hot-metal-wall vessel. Steam is generated to provide cooling for the pressure vessel while the fluid bed reaction takes place at temperatures as high as 2000 F. System pressure can be as low as 100 psia (minimum level is determined by economics), but in the combined-cycle application, the pressure would be determined by gas-turbine pressure ratio, pressure drop in the fuel-processing system, and gas-turbine fuel-control-system requirements.

The operating conditions within the gasifier result in a product gas free of tars and oils. Thus, no special cleanup procedures are needed.

Projected commercial-sized U-Gas gasifiers would handle 3000 ton/day of coal. It is estimated that the gasifier would have a 10-to-1 turndown ratio.

The Foster Wheeler Energy Corporation is currently designing an air-blown version of a two-stage, entrained-flow gasifier developed by Bituminous Coal Research (BCR). The oxygen-blown BCR gasifier is presently undergoing testing in a 120-ton/day pilot plant producing 2.5 million scf/day of syn gas at Homer City, Pennsylvania.

Run-of-the-mine coal is crushed, dried to two-percent moisture, and pulverized. Coal is metered from the feed hopper to an injector and then into hot transport gas (recycled from the gas-purification section) before being fed into the upper stage of the gasifier. In this stage, the coal reacts with synthesis gas from the lower stage and steam to produce methane, carbon monoxide, hydrogen, and unreacted char. The gases leave the upper stage at around 1800 F.

Entrained residual char is removed from the gas by cyclone separators and recycled via superheated steam to the lower stage of the gasifier. The char then reacts with steam and air at about 2800 F to form synthesis gas and molten slag. The hot synthesis gas flows to the upper stage for reaction with coal as described above. Molten slag collects and drains from the bottom of the lower stage into the slag pot, where it is water-quenched.

Overall, the upper-stage gasifier reactions are endothermic, and the process-heat requirement is supplied by combustion of char with air. The air rate is regulated to maintain the operating temperature in the upper stage, while the lower-stage temperature is controlled by steam addition. Temperature in the lower stage is fairly critical, because too high a temperature will damage the refractory and too low a temperature will cause the slag to freeze and accumulate.

While the pilot installation of the FW gasifier will have only a 480 ton/day capacity, it would appear that larger gasifiers of a capacity similar to the Texaco gasifier (1900-2000 ton/day) would be commercially viable. The operating pressure would be dependent upon the gas-turbine pressure ratio and pressure drop in the fuel-processing system and gas-turbine fuel controls.

The Texaco gasifier has been in commercial use for a number of years, producing synthesis gas (H_2 + CO) from a variety of liquid hydrocarbon feeds. Its application to coal gasification is still in the development stage, although as early as the mid-1950's a 100 ton/day pilot plant was operated at 300 psig on West Virginia coal. In Texaco's Montibello, California, research facility, a 15 ton/day pilot plant has provided low- and medium-Btu gas from coal to a gas-turbine combustor. Texaco and Southern California Edison are presently negotiating for a western-coal-fueled gasification demonstration plant which would provide medium-Btu gas to a utility boiler. This plant could eventually be converted to a combined-cycle installation.

In the Texaco gasifier, coal reacts with air and steam under slagging conditions in a refractory-lined pressure vessel. The hot gases leave the gasifier and pass through a fire-tube-type waste-heat-recovery steam generator. Slag is quenched at the bottom of the gasifier and removed via a lock hopper.

Pulverized coal (70% < 200 mesh) is slurried in water and pumped into the gasifier. The gasifier operates at temperatures above the ash fusion point; thus, no tars or oils are produced. Typically, the operating pressure would be 600 psig and above. A single nine-ft-diameter gasifier of the type commercially available would gasify approximately 1900 ton/day of coal.

Turndown to 50% of capacity is routinely accomplished in commercial applications and the gasifier will operate satisfactorily at 15% capacity.

2.3 FLUIDIZED-BED COMBUSTION (FBC)

2.3.1 Technology Description

The basic features of a coal-fired fluidized-bed boiler are depicted in Fig. 3.[9] Crushed coal (typically 1/4-in. top size) is continuously fed into and burned in a bed of crushed limestone or dolomite (typically 1/8-in. top size) that is fluidized by a continuous upward flow of air through a perforated plate at the bottom of the vessel. Boiler tubes typically are submerged in the bed and may also be situated in the freeboard region above the bed. SO_2 released from the burning coal reacts chemically with the limestone or dolomite in the bed, forming solid $CaSO_4$ and obviating the need for any subsequent flue-gas desulfurization. For most effective SO_2 sorption, the bed temperature is maintained in the range of 1500-1700°F. This relatively low combustion temperature has the additional advantages of reducing the formation of NO and the volatilization of trace elements, as well as precluding slag formation. Sorbent reactivity is maintained at the required level by continuously feeding fresh stone to the bed and withdrawing spent (sulfated) stone from the bed at equivalent rates. A major fraction of the coal ash is entrained and carried out of the bed with the combustion gases.

The flow of air and combustion bases upward thru the bed gives rise to a fairly rapid circulation and mixing of the bed solids. This action, in turn, results in very effective heat transfer to the submerged boiler tubes, nearly uniform temperature distribution throughout the bed, high volumetric heat-release rates, and the capability of burning nearly any type of solid fuel. The high rates of heat release and transfer also offer the promise that a fluidized-bed boiler can be smaller and more efficient than a conventional pulverized-coal boiler with the same power output.

For purposes of generating electric power, the energy-conversion efficiency can, in principle, be still further increased by pressurizing the fluidized-bed combustor (say to $\sim$ 10 atm) and generating additional power by expanding the hot, pressurized combustion gases through a gas turbine. Also,

BASIC FEATURES OF A FLUIDIZED-BED BOILER

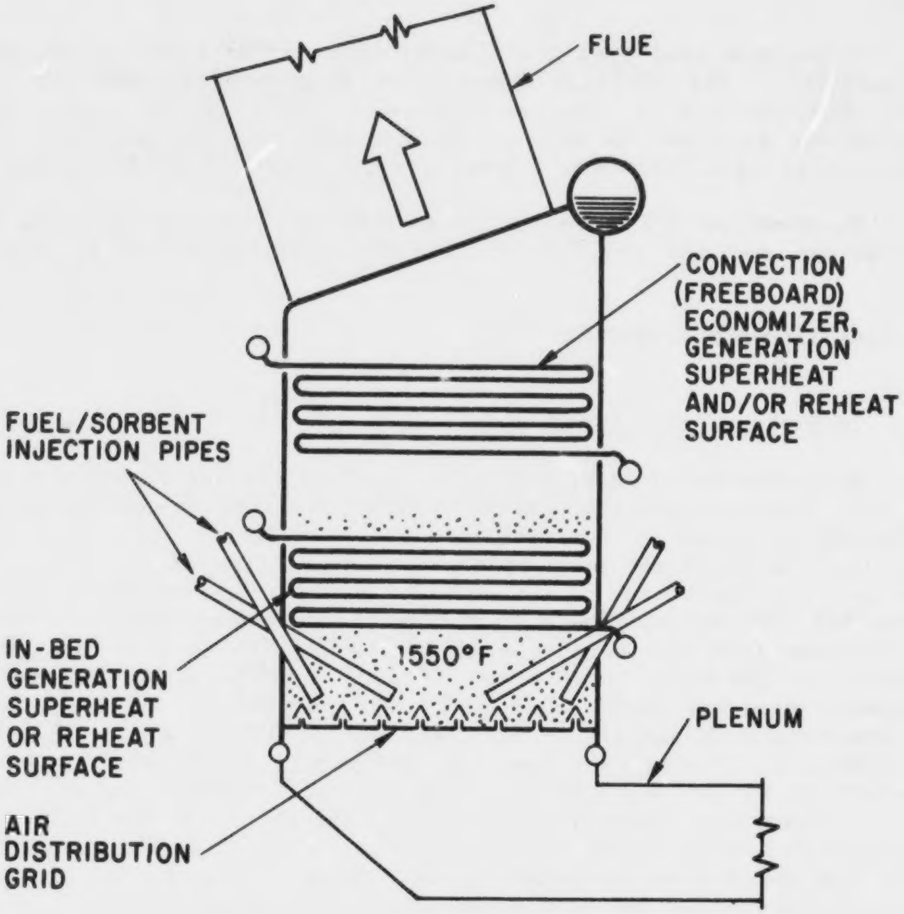

FLUE
CONVECTION (FREEBOARD) ECONOMIZER, GENERATION SUPERHEAT AND/OR REHEAT SURFACE
FUEL/SORBENT INJECTION PIPES
IN-BED GENERATION SUPERHEAT OR REHEAT SURFACE
1550°F
PLENUM
AIR DISTRIBUTION GRID

Figure 3

pressurized fluidized-bed combustion (PFBC), as opposed to atmospheric-pressure fluidized-bed combustion (AFBC), offers the following additional advantages: (1) more compact combustor vessels, (2) more effective SO_2 sorption (provided that dolomite or precalcined limestone is employed), (3) inherently lower NO emission, and (4) improved combustion efficiency. However, transport of solids into and out of the pressurized combustor and, to an even greater degree, protection of the gas turbine from excessive corrosion and erosion when driven by the relatively dirty combustion gases present formidable technical problems.

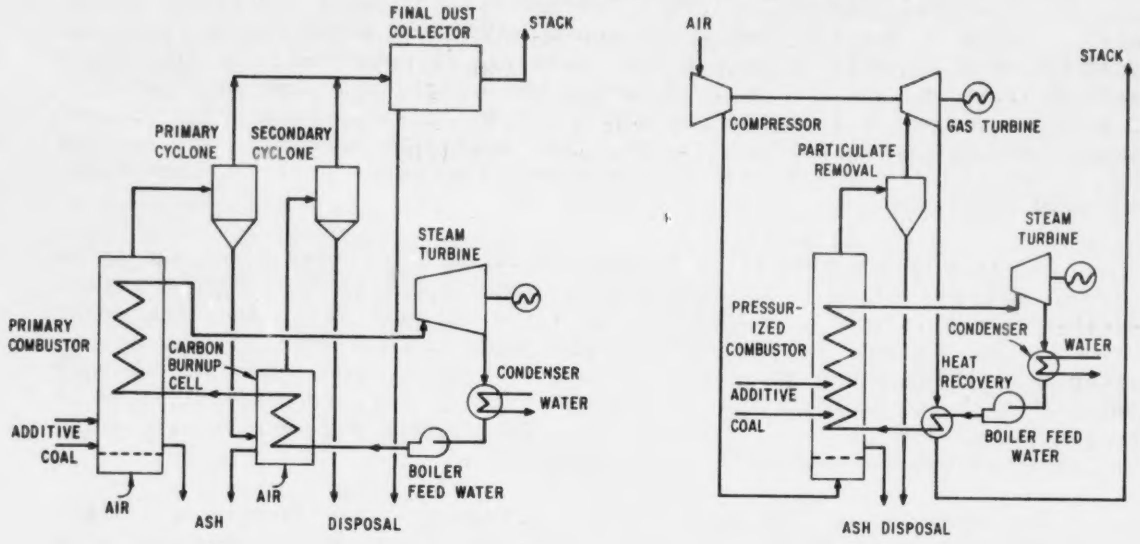
Figure 4 shows simplified schematics of an AFBC power plant and three different types of PFBC/combined-cycle power plants.[10] In this figure "additive" means sorbent (limestone or dolomite) and "ash" includes spent sorbent. In the case of the AFBC system, note the presence of the carbon-burnup cell (CBC) whose purpose is to recover the heating value of unburned coal particles and soot elutriated from the primary combustor and collected in the primary cyclone. It is believed that PFBC systems will not require a CBC to achieve acceptable combustion efficiency.

The water-cooled PFBC system shown in Figure 4b is essentially similar to the AFBC system in Figure 4a, except that the pressurized combustion gases are used to drive a gas turbine that might generate ∿ 20% of the electrical output of the plant. Both systems would be run with about 15-20% excess air.

The adiabatic PFBC shown in Figure 4d has no heat-transfer surface in the combustor. Instead, the combustor is supplied with ∿ 300% excess air, which serves to control the combustion temperature and carries off the heat released by the burning coal. In this case ∿ 80% of the plant electrical output would be generated by the gas turbine and ∿ 20% by a steam turbine connected to the waste-heat boiler. This configuration has the disadvantage that the particulate-removal equipment between the combustor and the gas turbine would have to handle very large volumes of hot, high-pressure gases. Also, there is concern that the large amount of excess air fed to the combustor would increase the emission of NO and particulates in the combustion gases.

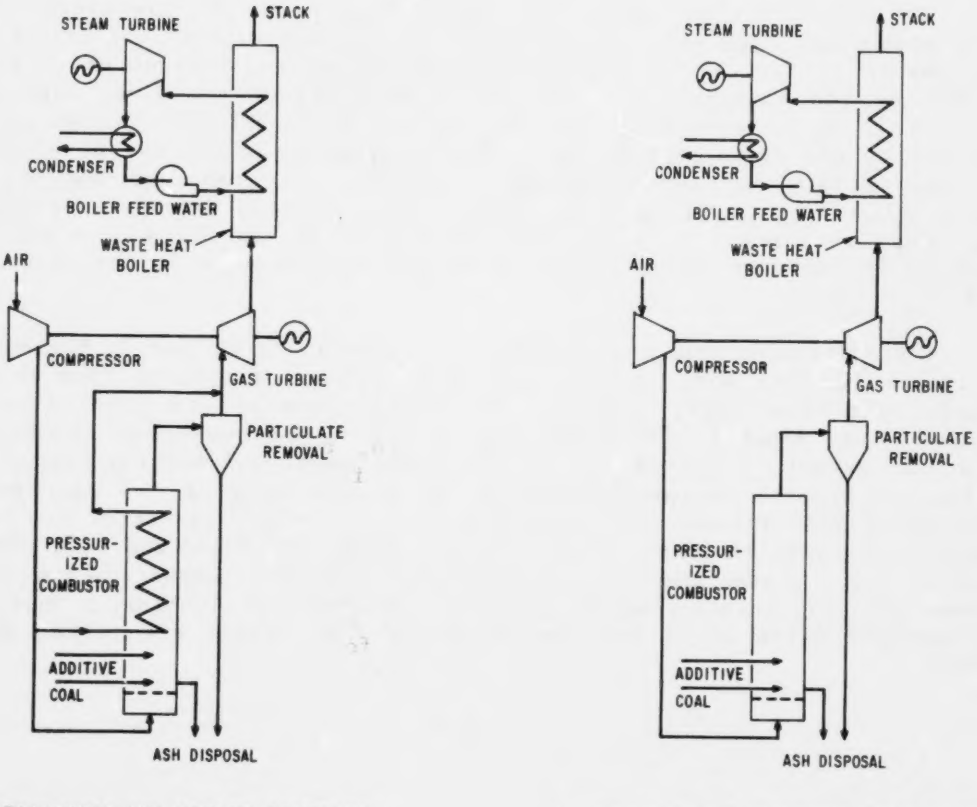
The air-cooled PFBC system shown in Figure 4c is similar to the adiabatic system in that most of the electrical output is derived from the gas turbine. In this case, however, the pressurized air is split into two streams. Only about 25-30% of the air is fed to the combustion chamber, so the actual amount of excess air and the quantity of gas passing through the particulate removal equipment would be no greater than in the case of the water-cooled PFBC system. The remaining 70-75% of the pressurized air passes through heat-transfer tubes located in and above the fluidized bed, and the resulting hot, pressurized air is conjoined with the combustion-gas stream between the particulate-removal equipment and the gas turbine. Thus, the disadvantages cited above for the adiabatic PFBC scheme are rather neatly avoided.

FLUIDIZED-BED COMBUSTION POWER GENERATION SYSTEMS


FINAL DUST COLLECTOR
STACK
PRIMARY CYCLONE
SECONDARY CYCLONE
STEAM TURBINE
PRIMARY COMBUSTOR
CARBON BURNUP CELL
CONDENSER
WATER
ADDITIVE
COAL
AIR
AIR
BOILER FEED WATER
ASH
DISPOSAL
AIR
STACK
COMPRESSOR
GAS TURBINE
PARTICULATE REMOVAL
STEAM TURBINE
PRESSUR-IZED COMBUSTOR
CONDENSER
WATER
HEAT RECOVERY
ADDITIVE
COAL
BOILER FEED WATER
ASH DISPOSAL

a. ATMOSPHERIC WATER-COOLED COMBUSTOR

b. PRESSURIZED WATER-COOLED COMBUSTOR


STEAM TURBINE
STACK
CONDENSER
BOILER FEED WATER
WASTE HEAT BOILER
AIR
COMPRESSOR
GAS TURBINE
PARTICULATE REMOVAL
PRESSUR-IZED COMBUSTOR
ADDITIVE
COAL
ASH DISPOSAL
STEAM TURBINE
STACK
CONDENSER
BOILER FEED WATER
WASTE HEAT BOILER
AIR
COMPRESSOR
GAS TURBINE
PARTICULATE REMOVAL
PRESSUR-IZED COMBUSTOR
ADDITIVE
COAL
ASH DISPOSAL

c. PRESSURIZED AIR-COOLED COMBUSTOR

d. PRESSURIZED ADIABATIC COMBUSTOR

Figure 4

2.3.2 Developmental Status and Prospects

In Table 4 are listed some of the more important FBC units that are presently in service or scheduled for early completion. For both AFBC and PFBC the list includes: (1) a pilot plant, (2) a components test and integration unit (CTIU) or equivalent test facility, and (3) a relatively small but presently operational process development unit (PDU).

The AFBC pilot plant at Rivesville, West Virginia, is located in a conventional, coal-fired power station of the Monongahela Power Co. The configuration consists of three main beds, each 10 x 12 ft in area, plus a slightly smaller bed that serves as a carbon-burnup cell. The unit can generate 300,000 lb of steam per hour at 1250 psig and 925 F. The steam can be fed to one of the existing turbogenerators at the station, producing electricity for the Allegheny Power System's commercial grid. The first commercial power generation with this unit took place in September 1977, and during May 1978 the unit was operated continuously for a period of 50 hours, producing 800 MWh of electricity. On August 9, 1978, a fire in the air preheater resulted in damages to the pilot plant estimated at $1.5 million. Repairs were expected to take about six months. Prior to the fire, a total of more than 70 hours of commercial operation had been logged.

The PFBC pilot plant being designed and built by Curtiss-Wright will be of the air-cooled, combined-cycle type shown schematically in Figure 2c and is expected to be operational by 1980. The combustor will feature a fluidized bed 12 ft in diameter with a height of 16 ft plus 15 ft of freeboard. Only 1/3 of the air from the compressor will be fed directly into the combustor; the remaining 2/3 will be heated by passage through tubes immersed in the bed. The gas turbine will generate 7 MW of electrical power. The exhaust gases will then pass through a waste-heat boiler, wherein steam will be produced at a rate equivalent to an additional 6 MW of electrical power. The gas-turbine blades will be of a special design that provides a "boundary layer" flow of cool, high-pressure air to protect the blades from corrosion and erosion.

There are, of course, certain unresolved issues and problems in connection with FBC power generation. Some of the more critical are listed in Table 5. Nevertheless, the present consensus seems to be that FBC power generation -- at least the atmospheric-pressure type -- can be brought to commercial status within a reasonably short period of time and that it will prove to be competitive with, or superior to, conventional PC/FGD power-generation technology with regard to cost, efficiency, and environmental acceptability.

Thus, a report recently drafted by a DOE task force concludes that AFBC power-generation technology is ready for commercialization and recommends that a 200-MWe commercial demonstration plant be constructed and put into operation by 1985. Conceptual design studies for such a plant have already been completed, but detailed design and construction would require about five years. Assuming that the demonstration plant is completed on schedule, the report projects that 60-120 GW of commercial AFBC utility-plant generating capacity could be on line by the year 2000.[11]

Table 4. Roster of Selected FBC Facilities

Installation	Type	Rated Power	Status (1/79)
Atmospheric-Pressure FBC Facilities			
1. Pope, Evans & Robbins, Inc. Rivesville, W. Va. (DOE)	Pilot plant	88 MW(t) 30 MW(e)	Temporarily out of service for repairs.
2. Morgantown Energy Technology Center Morgantown,W. Va. (DOE)	CTIU	18 MW(t)	Under construction; completion in 1980-81.
3. Babcock & Wilcox Co. Alliance, Ohio (EPRI)	PDU	6 MW(t)	Operational.
Pressurized FBC Facilities			
4. Curtiss-Wright Corp. Wood-Ridge, N.J. (DOE)	Pilot Plant	38 MW(t) 13 MW(e)	Under construction completion in 1980.
5. International Energy Agency Grimethorpe, England (IEA)	Flexible test facility	80 MW(t)	Under construction; startup later in 1979.
6. Exxon Research & Engineering Co. Linden, N.J. (EPA)	PDU ("Miniplant")	1.8 MW(t)	Operational.

ABBREVIATIONS: CTIU = Components Test and Integration Unit

PDU = Process Development Unit

Table 5. Key Development Issues in Fluidized-Bed Combustion

Issue	Relative Importance AFBC	Relative Importance PFBC
1. Sorbent Requirements and Utilization	****	***
2. Disposal or Utilization of Ash and Spent Sorbent	****	***
3. Combustion-Gas Cleanup at High Temperature and Pressure	N.A.	****
4. Particulate Removal from Stack Gas	***	*
5. Gas-Turbine-Blade Erosion, Corrosion and Deposition	N.A.	****
6. Solids Feeding and Transport Systems	***	****
7. Corrosion, Erosion, and Deposition in the Combustor	**	**
8. Improved Combustion Efficiency	***	**
9. Startup, Turndown, and Load-following Capability	**	***

NOTE: The number of asterisks indicates relative importance in each case. N.A. means "not applicable."

The TVA also has an active program looking toward the design and construction of AFBC power plants at the commercial scale. According to a recent report, their timetable calls for construction of a 200-MWe demonstration plant by 1984 and of a full-scale commercial plant by 1990.[12]

Although the prevailing view in the U.S. is that PFBC/combined-cycle power-generation technology is not yet ready for commercialization, at least one U.S. utility consortium seems to feel otherwise. After obtaining encouraging results during a feasibility study conducted jointly with Stal-Laval Turbine of Stockholm and Babcock & Wilcox, Ltd. of Birmingham, England, American Electric Power Co. has tentatively decided to proceed with design and construction of a commercial-scale demonstration plant, apparently of the water-cooled PFBC/combined-cycle type. The coal-fired plant, which AEP says might be in full operation as early as 1983, will be built at a decommissioned power plant in Brilliant, Ohio, and will utilize a 105-MWe steam turbogenerator already located at that station. Babcock & Wilcox will supply the pressurized combustor, while Stal-Laval will provide a suitable gas turbogenerator capable of generating 65 MWe. Details of the provisions for hot-gas cleanup and turbine-blade protection are not yet available.[13]

2.3.3 Environmental Implications of FBC

Table 6 presents, in outline form, a resume of the prospects for control of stack-gas emissions from coal-fired FBC power plants. Of the three pollutants covered by the federal New Source Performance Standards (NSPS), it would seem that particulates may pose the most serious control problems. First of all, particulate loadings in the combustion gases from an FBC are inherently rather high as a result of the absence of any slagging of the coal ash and of the unavoidable elutriation from the fluidized-bed of small particles of sorbent. Particulate collection by electrostatic precipitation (even "hot" ESP, as used at the Rivesville AFBC pilot plant) is unlikely to be very effective, owing to the unusually high electrical resistivity of the particulates. More likely, baghouses, following conventional cyclones, will be the preferred technology for AFBC power plants. In the case of PFBC/combined-cycle plants, the degree of particulate removal required to avoid undue erosion in the gas turbine will probably insure that the stack gases will meet particulate emission standards. If not, then a baghouse might still be required at the stack for additional collection of fine particulates.

At least in principle, the lower combustion temperatures employed in FBC could enhance the formation and emission of polycyclic organic compounds, including various compounds known to be mutagenic and/or carcinogenic. Analyses of FBC stack gases reported to date have not shown significant concentrations of such organic compounds, but further determinations of such materials (and of respirable particulates) in the stack gases from larger FBC units, such as the Rivesville pilot plant, should be given high priority.

Tightening of SO_2 emission standards may pose more severe problems in terms of solid-waste disposal for FBC power plants than for conventional plants using flue-gas desulfurization. It has been estimated that for an AFBC plant burning a typical high-sulfur coal, attainment of 90% sulfur retention

Table 6. FBC Emissions-Control Outlook

SO_2 EMISSIONS:

- Current or proposed NSPS are attainable with either AFBC or PFBC for practically any coal.
- Up to 1 ton of limestone or dolomite sorbent per ton of coal may be required in some situations.

NO_x EMISSIONS:

- Expected from AFBC: 0.4-0.6 $lb/10^6$ Btu.
- Expected from PFBC: 0.2-0.4 $lb/10^6$ Btu.
- Further reduction possible by employing staged combustion techniques.

TRACE-ELEMENT EMISSIONS: (No present or proposed NSPS)

- May be less than from conventional coal combustion, owing to lower temperature.
- Definitive data not yet available.

EMISSIONS OF POLYCYCLIC ORGANIC COMPOUNDS: (No present or proposed NSPS)

- May be more than from conventional coal combustion, owing to lower temperature.
- Definitive data not yet available.

PARTICULATE EMISSIONS:

- >99.0% collection efficiency required to meet present NSPS.
- >99.7% collection efficiency required to meet proposed NSPS.
- >99.9% collection efficiency required to protect gas-turbine blades in PFBC/combined-cycle applications.
- "Best-bet" collection technology:

    Cyclones plus baghouse for AFBC

    High-efficiency cyclones plus granular-bed or porous-ceramic filter for PFBC/combined-cycle.

---

would require approximately twice as much limestone (and produce twice as much spent sorbent) as attainment of the present SO_2 standard of 1.2 lb/10^6 Btu.[14] The resulting exacerbation of the spent-sorbent disposal problem could well delay the commercialization of FBC power generation or force the premature implementation of developing technology such as sorbent pretreatment or regeneration to reduce the production of solid waste.

3 DESIGN AND ECONOMIC PREMISES

In order that direct comparisons with flue-gas-desulfurization processes reported on by the Tennessee Valley Authority (TVA) may be made, design and economic premises compatible with those used by TVA have been employed.[15]

The base case for the conceptual design and detailed engineering cost studies is an approximately 500 MWe (net) new utility power plant burning Illinois No. 6 coal with a sulfur content (dry) of 3.86%. This coal has a moisture content of 12%, an ash content of 8.82%, and a higher heating value of 12,771 Btu/lb (dry). A detailed analysis of this coal is shown in Table 7.

Table 7. Characteristics of Illinois No. 6 Coal

Coal Property	Value
Rank	HVC Bituminous
Proximate Analysis (wt% as rec'vd)	
Moisture	12.00
Ash	8.82
Volatile Matter	31.41
Fixed Carbon	47.77
Ultimate Analysis (wt% dry)	
Carbon	69.52
Hydrogen	5.33
Nitrogen	1.25
Sulfur	3.86
Oxygen	10.02
Ash	10.02
Heating Value (Btu/lb dry)	
HHV	12771
LHV	12222
Free Swelling Index	4.5
Grindability Index (Hardgrove)	57.4
Initial Ash Fusion, (F)	2120-2240
Ratio of pyritic to organic sulfur	1.14

The projected operating life of the utility power plant is assumed to be 30 years, representing 127,500 hours of generating capacity. This is an average of 4,250 hours of operation per year. The projected load factor for the first year of operation is 0.8, or 7,000 hours of operation

3.1 EMISSION REGULATIONS

All of the alternative power plants have been designed to comply with the proposed New Source Performance Standards (NSPS). For particulate emissions, this is a 0.03-lb/10^6 Btu input. At this writing, the proposed EPA standard for sulfur-oxide emissions is an 85% removal (daily average) of sulfur in the coal as mined, with an emission ceiling of 1.2 lb/10^6 Btu and a floor of 0.2 lb/10^6 Btu. The 85% removal requirement has been used as a design factor for all of the alternative technologies. In the case of the atmospheric fluidized-bed combustor, a limestone sorbent was used with a three-to-one calcium-to-sulfur ratio. For the pressurized fluidized-bed combustor, dolomite was used as a sorbent with a two-to-one calcium-to-sulfur ratio. It has been confirmed that these mole ratios can remove 85% of the sulfur in tests performed at Argonne National Laboratory and the EXXON miniplant. The gasification/combined-cycle power systems achieve sulfur removal by using standard industrial H_2S removal techniques, such as the Stretford process with a Claus tailgas-cleanup system. Sulfur removal in the solvent-refined-coal process takes place during the hydrogenation of the coal that is dissolved in a coal-derived solvent. The sulfur, as hydrogen sulfide, is flashed off, separated from the recycle hydrogen, and converted to elemental sulfur.

3.2 PLANT LOCATION

A southern Illinois plant site has been chosen for estimating purposes. This would place the plant near several demand centers and in close proximity to extensive coal fields, thus minimizing transportation costs.

3.3 PROJECT SCHEDULE

A construction start date of 1981 and a plant start-up date of 1985 have been assumed for all of the technologies. Costs have been calculated on the basis of 1980 dollars and have been scaled based on the extrapolated average annual Chemical Engineering Cost Indices as shown in Figure 5.

3.4 ECONOMIC ASSUMPTIONS

3.4.1 Indirect Investment Charges

This area includes the materials and labor for equipment and installation and all costs (such as architect and engineering fees, contractor expenses, construction expenses, and in-house engineering) that are necessary for construction of a grass-roots power plant. The engineering design and contingency factors are based on the developmental status of the technology and experience with engineering projects.

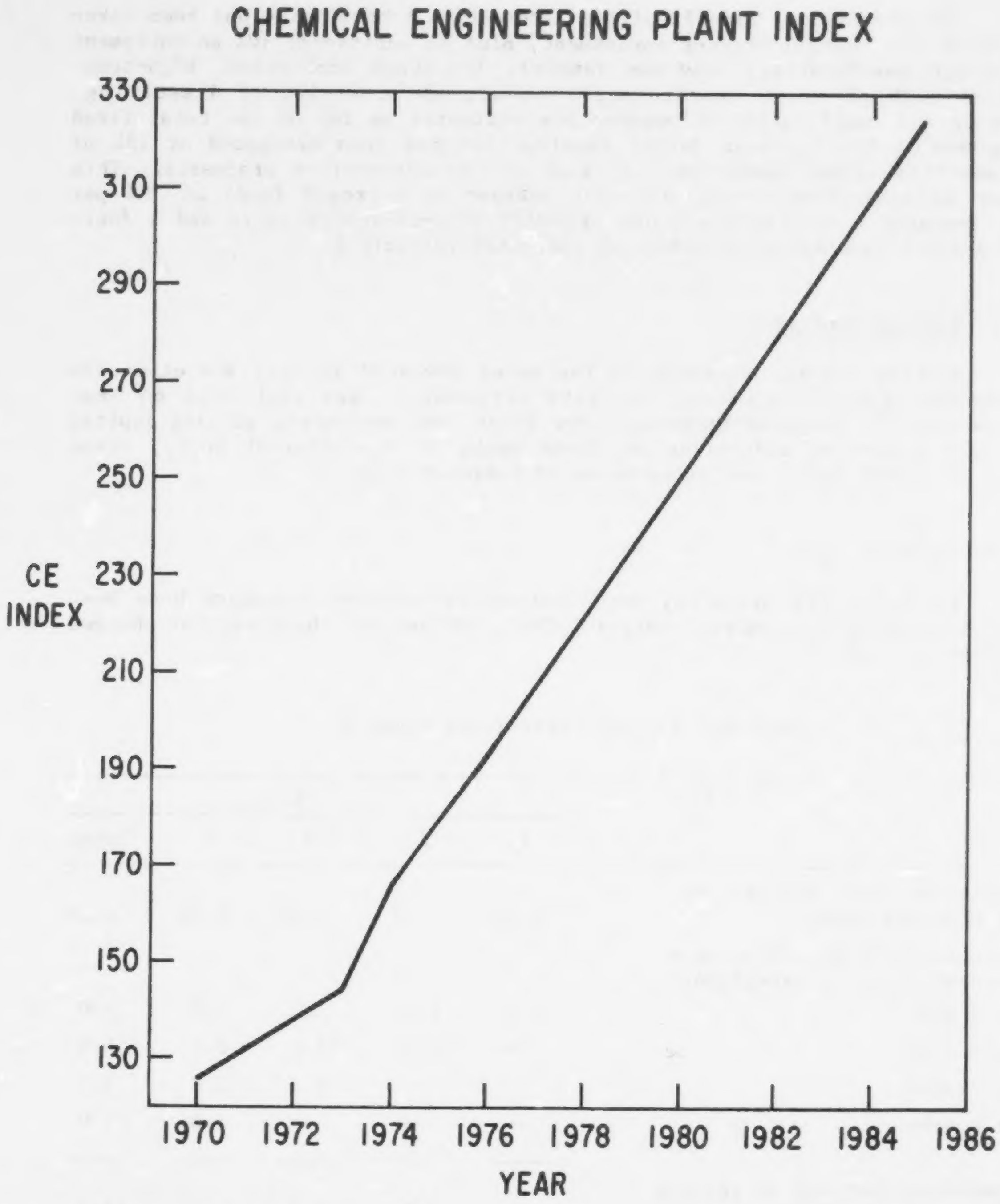

CHEMICAL ENGINEERING PLANT INDEX
CE INDEX
330
310
290
270
250
230
210
190
170
150
130
1970
1972
1974
1976
1978
1980
1982
1984
1986
YEAR

Figure 5

The contingency for all of the alternative technologies has been taken as 15% of the overall project investment, plus an additional 10% on equipment costs for gasification, acid-gas removal, fluidized combustion, high-temperature/high-pressure particulate removal, and coal-slurry dissolving. Start-up and modification allowances are estimated as 10% of the total fixed investment. The interest during construction has been estimated at 15% of the sub-total fixed investment for each of the alternative processes. This factor is equivalent to the interest accrued on borrowed funds at 10% per year assuming a capital structure of 60/40 debt-to-equity ratio and a four-year project expenditure schedule as indicated in Table 8.

3.4.2 Working Capital

Working capital consists of the money invested in coal and other raw materials carried in stock, accounts receivable, and cash kept on hand for payment of operating expenses. For these cost estimates, working capital has been taken as equivalent to three weeks of raw-material costs, seven weeks of direct costs, and seven weeks of overhead costs.

3.4.3 Indirect Costs

Following TVA practice, regulated-utility-company economics have been used in establishing capital charges. The breakdown for these capital charges is shown in Table 9.

Table 8. Project Expenditure Schedule

	Year				
	1	2	3	4	Total
Fraction of total expenditure as borrowed funds	0.10	0.20	0.20	0.10	0.60
Simple interest at 10%/yr as a percent of total expenditure					
Year 1 debt	1.0	1.0	1.0	1.0	4.0
Year 2 debt	--	2.0	2.0	2.0	6.0
Year 3 debt	--	--	2.0	2.0	4.0
Year 4 debt	--	--	--	1.0	1.0
Accumulated interest as percent of total expenditure	1.0	3.0	5.0	6.0	15.0

Table 9. Annual Capital Charges for Power Industry Financing

	Percentage of total depreciable capital investment		
	Years remaining life		
	30	25	20
Depreciation-straight line (based on years remaining life of power unit)	3.3	4.0	5.0
Interim replacements (equipment having less than 30-yr life)	0.7	0.4	-
Insurance	0.5	0.5	0.5
Property taxes	1.5	1.5	1.5
Total rate applied to original investment	6.0	6.4	7.0

	Percentage of unrecovered capital investment[a]
Cost of capital (capital structure assumed to be 60% debt and 40% equity)	
Bonds at 10% interest	6.0
Equity[b] at 14% return to stockholder	5.6
Income taxes (federal and state)[c]	5.6
Total rate applied to depreciation base	17.2[d]

a. Original investment yet to be recovered or "written off."

b. Contains retained earnings and dividends.

c. Since income taxes are approximately 50% of gross return, the amount of taxes is the same as the return on equity.

d. Applied on an average basis, the total annual percentage of original fixed investment for new (30-yr) plants would be 6.0% + 1/2 (17.2%) = 14.6%.

3.4.4 Overheads

Plant, administrative, and marketing overheads are costs that vary from company to company. With consideration of the various methods used in industry and illustrated in a variety of cost-estimating sources, the following method for estimating overheads is used.

Plant overhead is estimated as 50% of the subtotal conversion costs less utilities, and includes the projected costs for labor, maintenance, and analyses. Administrative overhead is estimated as 10% of operating labor and supervision. Marketing the product is considered in the estimation of overheads and is defined as 10% of sales revenue.

3.4.5 By-product Sales

In the evaluation of the annual revenue requirements, credit from sale of by-products (such as ammonia or sulfur recovered from low-Btu combined-cycle gasifiers) is deducted from the annual operating costs to obtain the net annual revenue requirements. The selling prices for sulfur and ammonia were taken as \$60 and \$100 per long ton, respectively.

3.4.6 Raw Material Costs

Costs for limestone and dolomite (needed for sulfur-oxides removal in fluidized-bed units) have been taken as \$10 per ton. For the purpose of calculating annual revenue requirements, a coal price of $\$1/10^6$ Btu has been assumed. This results in a net coal price of \$22.50 per ton of coal. In addition to this fixed coal price, a variation from $\$0.75-2/10^6$ Btu was also used, in order to study the sensitivity of busbar power costs to fuel costs.

4 ALTERNATIVE TECHNOLOGIES COMPARISON

Using the design and economic premises outlined in the previous section, a series of detailed capital and operating-cost estimates have been prepared for the technologies under discussion as alternatives to FGD. For fluidized-bed combustion, capital requirements were determined from detailed cost estimates prepared by Westinghouse Corporation for the Environmental Protection Agency.[16] The estimates for the four gasification/combined-cycle processes (Lurgi, IGT, Foster-Wheeler, and Texaco) were prepared based on unpublished information given to Argonne National Laboratory.[8] Information needed to determine capital requirements for both a solvent-refined-coal production plant and a power plant burning SRC were obtained from Reference 2.

4.1 HEAT RATES

The information noted above was used to determine heat rates for each of the alternative processes. These rates, which were used for capital design costing and revenue-requirement calculations, are shown in Table 10.

As can be seen, the heat rate for the pressurized fluidized-bed combustor is approximately 15% less than that of the atmospheric unit. This is due to two factors: 1) the increased pressures and temperatures of the

Table 10. Overall Heat Rates Of Alternative Processes

Process	Heat Rate (Btu/kWh)
AFBC	9618[a]
PFBC	8688[b]
G/CC:	
Lurgi	10856
IGT	8258
Foster Wheeler	8303
Texaco	8728
SRC	9000[c] 13040[d]

[a] 3:1 calcium/sulfur ratio-limestone sorbent.
[b] 2:1 calcium/sulfur ratio-dolomite sorbent.
[c] Based on SRC-burning power plant.
[d] Based on Btu content of coal input to SRC production plant.

pressurized FBC over that of the atmospheric FBC result in inherently more efficient operation, and 2) the ability of the pressurized fluidized-bed unit to utilize a 2-to-1 calcium-to-sulfur absorbent ratio as compared to the 3-to-1 ratio for the atmospheric combustor, results in less energy being used in heating and calcining of sorbent. The overall heat rates of the IGT, Foster-Wheeler, and Texaco gasifiers are less than that of the older Lurgi gasifier. Although differing by approximately 6%, the heat rates of these three gasifiers may be considered to be equivalent at this stage of development. For solvent-refined coal, the heat rate for a power plant utilizing SRC as a fuel is shown as 9,000 Btu/kWh. What is interesting to note, however, is that when one takes into account the Btu losses incurred during the SRC production process and relates the power-plant heat rate back to the inlet feed coal, a new heat rate of 13,040 Btu/kWh is found.

4.2 CAPITAL COSTS

4.2.1 Solvent-Refined Coal

The capital-cost breakdown for an SRC-I production plant is shown in Table 11. This cost estimate is based on a plant with a coal input of 20,000 ton/day, which is the size considered most likely for a commercial facility at this time. As a basis for comparison, it should be noted that such a plant operating at full capacity would produce enough fuel for a 1400-MW power plant. The capital costs include the equipment necessary for ash removal (filtration), sulfur removal, and sulfur recovery (as elemental sulfur).

4.2.2 Gasification/Combined-Cycle Power Plants

The capital costs for the four gasification/combined-cycle power plants analyzed in this paper are shown in Table 12. Depending upon the gasifier chosen, total capital investment can be seen to range from slightly over $450 million for the IGT unit to well over $600 million for the Lurgi-based power plant. All of the plants include facilities for acid gas (H_2S) removal and elemental sulfur recovery. There are also ammonia-recovery facilities (except for the Texaco gasifier), because this is a by-product of the gasification process.

4.2.3 Fluidized-Bed Combustors

The capital-cost breakdown for the atmospheric and pressurized fluidized-bed power plants are shown in Tables 13 and 14, respectively. These estimates include the equipment necessary for sorbent storage, pulverization, and feed to the fluidized-bed unit, as well as spent-sorbent removal.

4.3 ANNUAL REVENUE REQUIREMENTS

The annual revenue requirements for all of the alternative technologies are shown in Tables 15 through 22. In all cases except that of the SRC plant.

Table 11. Solvent-Refined-Coal Production Plant Capital-Cost Breakdown[a]

Categories	Costs (Millions of 1980 Dollars)
Coal Preparation	40.4
Coal-Slurry Dissolving, Benfield, Cryofining, and Oil Absorption	205.3
Filtration	160.4
Solvent Degassing and Recovery	66.0
Hydrogen Plant	
Koppers-Totzek	96.2
Air Separation	37.9
Acid-Gas Removal	29.8
Shift and Purification	25.1
Sulfur Recovery	18.4
Waste-Water Treatment	17.1
Product Storage & Shipping	9.4
Support Facilities	29.4
Total Plant Investment	735.4
A/E Home Office and Fee (10% of Estimated BOP)	44.1
Contingency[b]	143.4
Interest During Construction	138.4
Total Depreciable Investment	1061.3
Start-up and Modifications	106.1
Land	1.5
Working Capital[c]	47.9
Total Capital Investment	1216.8

[a] Approximately 20,000 ton/day coal input.

[b] 15% of plant investment plus 10% additional each on Coal-Slurry Dissolving, Benfield, Gasifier and Acid-Gas Removal.

[c] Equivalent to 3 weeks of raw materials, 7 weeks of direct costs, and 7 weeks of overhead costs.

Table 12. Gasification/Combined-Cycle Power Plant Capital-Cost Breakdown

Categories	Costs (Millions of 1980 Dollars)			
	Lurgi	IGT	Foster Wheeler	Texaco
Coal Handling	12.77	12.60	18.47	12.14
Oxidant Feed	2.47	2.58	2.49	4.01
Gasification and Ash Handling	63.01	22.50	27.42	33.08
Gas Cooling	29.97	44.05	25.73	73.99
Acid-Gas Removal and Sulfur Recovery	38.73	32.29	29.13	26.68
Process-Condensate Treatment	55.98	9.28	7.55	-----
Steam, Condensate and BFW	30.32	1.35	1.26	0.52
Support Facilities	25.50	23.45	22.48	15.95
Combined-Cycle Components	126.91	161.29	152.01	122.79
Total	385.66	309.39	286.54	289.16
A/E Home Office & Fee @ 10% of estimated BOP	25.27	20.28	18.78	18.95
Labor, Materials & Indirects Contingency	68.02	51.89	48.64	49.35
Interest during Construction	71.84	57.23	53.09	53.62
Total Depreciable Investment	550.79	438.79	407.05	411.08
Start up and Modifications	55.0	44.0	41.0	41.0
Land	1.5	1.5	1.5	1.5
Working Capital	17.60	13.79	12.82	12.86
Total Capital Investment	624.89	454.08	462.37	466.44

Table 13. AFBC Power Plant Capital-Cost Breakdown

Categories	Costs (Millions of 1980 Dollars) Major Components	1. Direct Labor	2. Indirect Field	3. Balance-of-Plant Materials	Total
Steam Generators	47.36	13.59	12.23	5.896	79.08
Turbine Generator	27.23	2.04	1.84	0.094	31.20
Process Mechanical Equipment	12.40	8.24	7.42	27.52	55.58
Electrical		16.48	14.83	11.51	42.82
Civil and Structural		13.89	12.50	12.83	39.22
Process Piping and Instrumentation		11.07	9.96	9.455	30.49
Yardwork and Miscellaneous		2.11	1.90	1.591	5.6
	86.99	67.42	60.68	68.91	283.99
	BOP Labor, Materials & Indirects (1 + 2 + 3)				197.01
	A/E Home Office & Fee @ 10%				19.70
	Contingency				50.29
	Interest during Construction				53.1
	Total Depreciable Investment				407.08
	Start-up and modifications				40.7
Land					2.4
Working Capital					14.06
Total Capital Investment					464.24

Table 14. PFBC Power Plant Capital-Cost Breakdown

Categories	Costs (Millions of 1980 Dollars) Major Components	1. Direct Labor	2. Indirect Field	3. Balance-of-Plant Materials	Total
PFB Steam Generators	41.35	5.32	4.79	2.655	54.11
Turbine Generators	47.44	2.09	1.88	.171	51.58
Process Mechanical Equipment	17.98	7.67	6.90	22.436	54.99
Electrical		11.64	10.48	9.42	31.54
Civil and Structural		12.21	10.99	9.59	32.79
Process Piping and Instrumentation		16.51	14.86	17.21	48.58
Yardwork and Miscellaneous		1.95	1.75	1.46	5.16
	106.77	57.39	51.65	62.94	278.75
	BOP Labor, Materials & Indirects (1 + 2 + 3)				171.98
	A/E Home Office & Fee @ 10%				17.20
	Contingency				51.45
	Interest during Construction				52.11
	Total Depreciable Investment				399.51
	Start-up and Modifications				40.0
Land					2.4
Working Capital					13.12
Total Capital Investment					455.03

Table 15. Total Average Annual Revenue Requirements - SRC Production Plant

	8,760 hr/yr (1.0 Capacity)	7,000 hr/yr (0.8 Capacity)
	Millions of 1980 Dollars per year	
Direct Costs		
Coal ($1/$10^6$ Btu)	194.0	155.2
Filter Aid and Other Chemicals	6.8	5.4
Raw Water ($.4/1000 gal)	1.9	1.5
Ash Disposal ($1/ton)	1.1	0.9
Electric Power (2.5¢/kWh)	35.0	28.0
Maintenance and Operation	24.8	22.3
Total Direct Costs	263.6	213.3
Indirect Costs		
Capital Charges		
Depreciation, replacements and insurance (6% of TDI)	63.7	63.7
Average cost of capital and insurance (8.6% of TCI)	104.6	104.6
Overheads		
Plant (50% of O&M)	12.4	11.2
Administrative (10% of O&M labor)	1.2	1.2
Total Indirect Costs	181.9	180.7
Total Costs	445.5	394.0
Hydrocarbon byproduct credit ($2/$10^6$ Btu)	(36.1)	(28.9)
Total Annual Revenue Requirements	409.4	365.1
Unit SRC Revenue Requirements $/$10^6$ Btu SRC output	3.13	3.42
$/ton SRC output	97.3	108.
$/$10^6$ Btu Coal input	2.11	2.35
$/ton Coal input	47.5	52.9

Table 16. Total Average Annual Revenue Requirements - SRC Power Plant[a]

	Annual Cost($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
SRC ($\$3.42/10^6$ Btu)	107,730,000	65,408,000
Maintenance and Operation	4,200,000	3,500,000
Total Direct Costs	111,930,000	908,000
Indirect Costs		
Capital Charges[b]		
Depreciation, interim replacements and insurance at 6% of total depreciable investment	18,000,000	18,000,000
Average cost of capital and taxes at 8.6% of total capital investment	27,090,000	27,090,000
Overheads		
Plant, 50% of O&M	2,100,000	1,750,000
Administrative, 10% of operating labor	100,000	100,000
Total Indirect Costs	47,290,000	46,940,000
Total Annual Revenue Requirements	159,220,000	115,848,000
Unit Revenue Requirements mills/kWh[c]	45.5	54.5
\$/ton coal[c]	78.1	93.7
$\$/10^6$ Btu input[c]	3.46	4.16

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[a] 500 MWe output.

[b] Capital charges based on \$600/kW depreciable investment and \$630/kW total investment.

[c] Based on coal feed to SRC production plant. 31% Btu loss in SRC production.

Table 17. Total Average Annual Revenue Requirements - Lurgi Combined-Cycle Plant[a]

	Annual Cost ($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
Coal ($1/$10^6$ Btu)	40,576,430	24,853,063
Catalyst and Chemicals	625,872	383,347
Utilities	732,024	448,365
Ash Disposal (on-site)	636,239	389,696
Maintenance and Operation	9,314,600	6,412,380
Sulfur Credit ($60/long ton)	(3,227,424)	(1,976,798)
Ammonia Credit ($100/long ton)	(1,859,138)	(1,138,722)
Total Direct Costs	46,798,603	29,371,331
Indirect Costs		
Capital Charges		
Depreciation, interim replacements and insurance at 6% of total depreciable investment	33,047,400	33,047,400
Average cost of capital and taxes at 8.6% of total capital investment	53,740,540	53,740,540
Overheads		
Plant, 50% of O & M	3,744,800	2,293,690
Administrative, 10% of operating labor	182,500	182,500
Total Indirect Costs	90,715,240	89,264,130
Total Annual Revenue Requirements	137,513,840	118,635,460
Unit Revenue Requirements mills/kWh	36.9	52.4
$/ton coal	76.3	107.
$/$10^6$ Btu input	3.39	4.77

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[a] 532.8 MWe output.

Table 18. Total Average Annual Revenue Requirements - IGT Combined-Cycle Plant[a]

	Annual Cost ($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
Coal ($1/$10^6$ Btu)	30,309,250	18,564,416
Catalyst and Chemicals	159,843	97,904
Utilities	440,000	269,500
Ash Disposal (on site)	477,404	292,410
Maintenance and Operation	6,933,320	4,924,784
Sulfur Credit ($60/long ton)	(2,608,433)	(1,597,665)
Ammonia Credit ($100/long ton)	(124,299)	(76,133)
Total Direct Costs	35,587,085	22,475,216
Indirect Costs		
Capital Charges		
Depreciation, interim replacements and insurance at 6% of total depreciable investment	26,327,400	26,327,400
Average cost of capital and taxes at 8.6% of total capital investment	39,050,880	39,050,880
Overheads		
Plant, 50% of O & M	3,591,660	1,587,392
Administrative, 10% of operating labor	175,000	175,000
Total Indirect Costs	69,144,940	67,140,672
Total Annual Revenue Requirements	104,732,025	89,615,888
Unit Revenue Requirements mills/kWh	28.6	40.3
$/ton coal	77.8	109.
$/$10^6$ Btu input	3,46	4.83

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[a] 523.2 MWe output.

	Annual Cost ($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
Coal ($1/$10^6$ Btu)	28,429,704	17,413,194
Catalyst and Chemicals	90,945	55,704
Utilities	628,571	385,000
Ash Disposal (on-site)	447,800	274,277
Maintenance and Operation	6,594,040	4,716,975
Sulfur Credit ($60/long ton)	(2,377,757)	(1,456,376)
Ammonia Credit ($100/long ton)	(1,315,232)	(805,580)
Total Direct Costs	32,498,071	20,583,194
Indirect Costs		
Capital Charges		
Depreciation, interim replacements and insurance at 6% of total depreciable investment	24,423,000	24,423,000
Average cost of capital and taxes at 8.6% of total capital investment	39,763,820	39,763,820
Overheads		
Plant, 50% of O & M	2,422,020	1,483,488
Administrative, 10% of operating labor	175,000	175,000
Total Indirect Costs	66,783,840	65,845,308
Total Annual Revenue Requirements	99,281,911	86,428,502
Unit Revenue Requirements mills/kWh	29.1	41.7
$/ton coal	78.6	112.
$/$10^6$ Btu input	3.49	4.96

---

[a] 488.1 MWe output.

Table 20. Total Average Annual Revenue Requirements - Texaco Combined-Cycle Plant[a]

	Annual Cost($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
Coal ($\$1/10^6$ Btu)	27,015,314	16,546,881
Catalyst and Chemicals	51,918	31,800
Utilities	847,262	518,948
Ash Disposal (on-site)	425,521	260,632
Maintenance and Operation	7,045,160	4,944,848
Sulfur Credit ($60/long ton)	(2,191,024)	(1,342,002)
Ammonia Credit ($100/long ton)	---------	---------
Total Direct Costs	33,194,151	20,961,107
Indirect Costs		
Capital Charges		
Depreciation, interim replacements and insurance at 6% of total depreciable investment	24,664,800	24,664,800
Average cost of capital and taxes at 8.6% of total capital investment	40,113,840	40,113,840
Overheads		
Plant, 50% of O & M	4,335,080	3,284,924
Administrative, 10% of operating labor	162,500	162,500
Total Indirect Costs	69,276,220	68,226,064
Total Annual Revenue Requirements	102,470,371	89,187,171
Unit Revenue Requirements mills/kWh	33.2	47.6
$/ton coal	85.3	121.
$\$/10^6$ Btu input	3.79	5.39

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[a] 441.2 MWe output.

Table 21. Total Average Annual Revenue Requirements - AFBC Power Plant[a]

	Annual Cost ($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
Coal ($\$1/10^6$ Btu)	33,673,500	20,444,625
Sorbent ($10/ton)	6,015,800	3,652,450
Spent-Sorbent Disposal (on-site)	2,523,920	1,532,380
Maintenance and Operation	6,910,000	5,590,000
Total Direct Costs	49,123,220	31,219,455
Indirect Costs		
Capital Charges		
Depreciation, interim replacements, and insurance at 6% of total depreciable investment	24,424,800	24,424,800
Average cost of capital and taxes at 8.6% of total capital investment	39,924,640	39,924,640
Overheads		
Plant, 50% of O&M	3,455,000	2,795,000
Administrative, 10% of operating labor	250,000	250,000
Total Indirect Costs	68,054,440	67,394,440
Total Annual Revenue Requirements	117,177,660	98,613,895
Unit Revenue Requirements mills/kWh	33.5	46.4
$/ton coal	78.3	109.
$\$/10^6$ Btu input	3.48	4.82

[a] 500 MWe output.

Table 22. Total Average Annual Revenue Requirements - PFBC Power Plant[a]

	Annual Cost ($)	
	7,000 hr/yr (0.8 Capacity)	4250 hr/yr (0.49 Capacity)
Direct Costs		
Coal ($1/$10^6$ Btu)	30,397,500	18,455,625
Sorbent ($10/ton)	6,136,200	3,725,550
Spent-Sorbent Disposal (on-site)	2,287,040	1,388,560
Maintenance and Operation	7,070,000	5,730,000
Total Direct Costs	45,890,740	29,299,735
Indirect Costs		
Capital Charges		
Depreciation, interim replacements, and insurance at 6% of total depreciable investment	23,970,600	23,970,600
Average cost of capital and taxes at 8.6% of total capital investment	39,132,580	39,132,580
Overheads		
Plant, 50% of O&M	3,535,000	2,865,000
Administrative, 10% of operating labor	250,000	250,000
Total Indirect Costs	66,888,180	66,218,180
Total Annual Revenue Requirements	112,778,920	95,517,915
Unit Revenue Requirements mills/kWh	32.2	45.0
$/ton coal	83.5	116.
$/$10^6$Btu input	3.71	5.18

[a] 500 MWe output

revenue requirements for both 7,000 hours a year (0.8 capacity) and 4,250 hours per year (0.49 capacity) have been shown. In the case of the SRC production plant, the annual revenue requirements necessary for full capacity (8,760 hours per year) and 0.8 capacity (7,000 hours per year) have been shown. The 7,000 hours of operation per year is considered the norm at which the SRC plant will be operating. Also, unit revenue requirements are shown both in terms of SRC fuel output and in terms of coal input. It should be noted that almost half of the total annual revenue requirement for the SRC production plant is made up of the cost of coal needed to produce the SRC. This large amount of coal is necessary to produce the required amount of SRC-I due in part to the process thermal efficiency. In a report by Air Products and Chemicals, Inc. and Catalytic, Inc.[2] it was reported that "A direct ratio of saleable product to total coal feed gives a plant thermal efficiency of 78%." This is based on treatment of the plant's electrical requirements as an operating rather than an energy expense. If electricity use is included in the energy balance, the ratio of saleable product to total energy input gives a plant thermal efficiency of 73%.

Table 16 utilizes a fuel cost of $\$3.42/10^6$ Btu for determination of the annual revenue requirement for a power plant burning SRC. It should be noted, in this case, that two-thirds of the annual-revenue requirement consists of costs expended for SRC fuel itself and that the unit revenue requirement (mills/kWh) for the SRC power plant is the highest of any of the alternative technologies studied.

The average annual revenue requirements for the gasifier processes covered in this report can be seen in Tables 17 through 20. It was found that the IGT and Foster-Wheeler gasifiers result in nearly identical busbar-power costs. These two alternative technologies also had the lowest unit revenue requirements of any looked at in this study. The Texaco gasifier was found to have a unit revenue requirement greater than that of either of the previous two, while the Lurgi gasifier had the highest unit revenue requirement of any of the gasification/combined-cycle power systems. It should be noted that in all cases (where applicable) both sulfur credits and ammonia credits for sale of by-products have been factored into the revenue calculations.

Tables 21 and 22 display the revenue requirements necessary for operation of atmospheric and pressurized fluid-bed power plants, respectively. The analysis found that the pressurized fluid bed has a smaller unit revenue requirement than the atmospheric. This is due in part to the higher efficiency of the pressurized fluid-bed unit, resulting in lower capital costs, as previously discussed, and lower operating costs, i.e., less coal consumed.

5 CONCLUSIONS

A summary of the operating and economic characteristics of the alternative processes is shown in Table 23. The overall heat rates of the pressurized-fluid-bed, IGT gasifier, Foster-Wheeler gasifier, and Texaco gasifier power plants are approximately equivalent and are lower than that of comparable conventional coal-burning power plants with flue-gas desulfurization (about 9,900 Btu/kWh). The total capital investment, in terms of dollars per kW, would tend to indicate that the SRC power plant is the cheapest one to construct. This cost of $630 per kW is less than for even a conventional power plant, in 1980 dollars. It should be noted, however, that the total capital investment for the SRC power plant does not include that for the SRC fuel-production facility. If this cost is included as part of the total capital investment, then greater than $1,100 of investment per kW is needed. This is comparable to the investment needed for a Lurgi combined-cycle plant, which was the most expensive of the units studied. Examination of the total power cost, in mills/kWh, of the alternative processes leads to several conclusions, including: 1) at the current stage of development, solvent-refined coal is not a viable option for replacement of conventional coal power generation with flue-gas desulfurization; 2) of the gasification/combined-cycle systems studied, the IGT and Foster-Wheeler processes offer the best possibilities for replacement of conventional coal-burning power plants (Whether or not the Texaco gasification process will be able to compete successfully with these will depend upon further cost refinements and reduction in overall capital requirements.), and 3) pressurized fluidized-bed combustion is a viable option for electric-power generation and may become even more so if sorbent-regeneration processes are successfully developed.

To further investigate tradeoffs between the technologies, the annual unit revenue requirements are plotted as a function of fuel cost in Figure 6. A change in the cost of coal is shown to have little or no effect on the economic rankings, although some of the processes show a greater dependence on fuel cost than do others.

In summary, seven processes have been examined as possible alternatives to conventional pulverized-coal combustion with flue-gas desulfurization. (Power costs for such a system are estimated at 34 mills/kWh for this comparison.) SRC is clearly the highest-cost alternative, at over 45 mills/kWh, with the Lurgi-based G/CC system second at about 37 mills/kWh. The remaining five options all fall between approximately 29 and 33 mills/kWh, a variation of about 13%, which is well within the uncertainties inherent in costs of developing technologies. The lowest-cost options in this group are the G/CC plants utilizing the IGT and Foster-Wheeler gasifiers. These are followed by the PFBC plant, which is also a combined-cycle system. This type of cycle helps to increase plant efficiency, but at the expense of a more complex system.

It should be noted in closing that all of the processes looked at under this study are at best in the pilot-plant or demonstration stage insofar as electric-power generation is concerned. Therefore, great caution should be taken in using the costs given in this paper as absolute rather than relative numbers. Problems with scale-up, discoveries during development, and changes in regulatory constraints can deeply affect the results that have been shown here.

Table 23. Alternative Processes Summary

			Gasification Combined-Cycle Systems				
	AFBC	PFBC	LURGI	IGT	F-W	TEXACO	SRC
MWe net	500	500	532.8	523.2	488.1	441.2	500
Overall Heat Rate (Btu/kWh)	9618	8688	10856	8258	8303	8728	9000 (13040)
Total Capital Investment ($/kW)	928	910	1173	868	947	1057	630
Capital Cost[a] (mills/kWh)	18.4	18.0	23.3	17.9	18.8	21.0	12.9
Fuel cost[b] (mills/kWh)	9.6	8.7	10.9	8.3	8.3	8.7	30.8
Other Costs (mills/kWh)	5.5	5.5	2.7	2.4	2.0	3.5	1.8
Total Power Cost[c] (mills/kWh)	33.5	32.2	36.9	28.6	29.1	33.2	45.5

[a] Based on 6% of Total Depreciable Investment and 8.6% of Total Capital Investment.

[b] Based on $1.00/MBtu base coal cost.

[c] First-year costs for 7000 hours of operation. The corresponding cost for a conventional power plant equipped with FGD is estimated to be 34 mills/kWh.

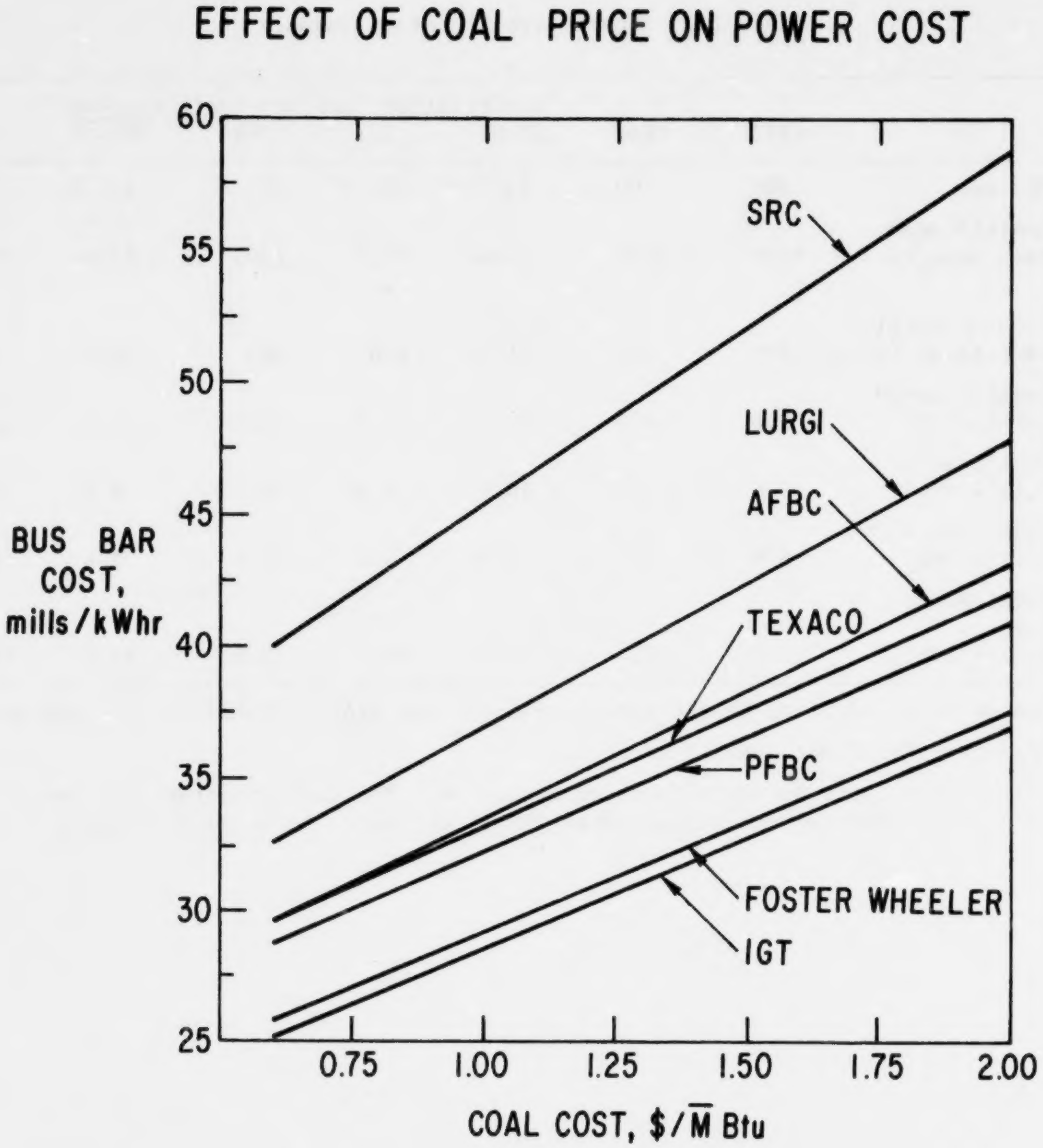
EFFECT OF COAL PRICE ON POWER COST
BUS BAR COST, mills/kWhr
60
55
50
45
40
35
30
25
0.75
1.00
1.25
1.50
1.75
2.00
COAL COST, $/$\overline{M}$ Btu
SRC
LURGI
AFBC
TEXACO
PFBC
FOSTER WHEELER
IGT

Figure 6

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ECONOMICS AND ENERGY REQUIREMENTS OF SULFUR OXIDES CONTROL PROCESSES

By

G. G. McGlamery, T. W. Tarkington, and S. V. Tomlinson
Emission Control Development Projects
Tennessee Valley Authority
Muscle Shoals, Alabama

Prepared for Presentation at
Flue Gas Desulfurization Symposium
Sponsored by U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Research Triangle Park, North Carolina
Held at
Caesars Palace
Las Vegas, Nevada
March 5-8, 1979

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ABSTRACT

As part of a continuing program to evaluate the design, energy consumption, and economics of sulfur oxides control processes, this paper presents the results from three separate studies being carried out for EPA by TVA. Energy and preliminary economic requirements for three physical and three chemical coal-cleaning processes are given along with similar evaluations of a number of FGD processes. The FGD evaluations cover technical updates of older systems such as the lime and limestone throwaway processes and the magnesia and sodium scrubbing to produce sulfuric acid. In addition, a sulfur-producing option, Wellman-Lord scrubbing with coal reduction of SO_2 by Allied Chemical, is presented for the first time.

ECONOMICS AND ENERGY REQUIREMENTS OF SULFUR OXIDES CONTROL PROCESSES

INTRODUCTION

The EPA-TVA interagency program to evaluate the design and economics of sulfur oxides control processes for fossil-fueled power plants continues into its twelfth year. Nine major reports have been distributed, three are about to be published, and projects which will result in five more are now underway. In addition to conceptual design and comparative economic evaluation of FGD processes, these projects include byproduct marketing studies, sludge disposal economics, and an energy and economic evaluation of physical and chemical coal-cleaning techniques. Throughout this period, research groups, utilities, vendors, and regulatory agencies have utilized these studies in decision-making processes related to sulfur oxides control.

In this paper the preliminary energy requirements and economic results of three EPA-sponsored studies are presented. Data are reported from a study of physical and chemical coal-cleaning processes. A ground-to-ground energy study of limestone, lime, and magnesia scrubbing FGD economics is discussed, with emphasis on updated technology for the magnesia process. Also discussed is a second FGD evaluation of the Wellman-Lord FGD process technology coupled with either sulfuric acid or sulfur production using the Allied Chemical coal-reduction process. These two FGD studies are evaluated on the same basis and are also comparable to the limestone, double-alkali, and citrate process economics presented at the November 1977 FGD Symposium by R. L. Torstrick, et al., Economic Evaluation Techniques, Results, and Computer Modeling for Flue Gas Desulfurization. (Proceedings: Symposium on Flue Gas Desulfurization, Volume II; EPA 600/7-78-058b, 1978)

The premises and cost values used in these companion FGD evaluations are listed in Appendix A. Slightly different premises used in the coal-cleaning study also are shown in Appendix A.

In the future, changing conditions will require modifications to many of the premises used in these studies. Emission regulations, economic conditions, and industry practices are all changing. Premises covering equipment redundancy; particulate, SO_x, and NO_x removal; waste disposal techniques; flue gas bypassing; reduced on-stream time for boilers; and allowance for inflation over the operating life of the system are some which will be revised in the near future.

PHYSICAL AND CHEMICAL COAL CLEANING

During recent years there have been indications that physical coal cleaning (PCC) and chemical coal cleaning (CCC), either separately or in combination with FGD, could help utilities reduce the cost of meeting sulfur oxides emission requirements. At EPA's request a study was undertaken by TVA to evaluate the economics of three PCC processes and three CCC processes. Preliminary results of this study presented here are part of a larger study to be published later this year. In addition to the results described in this paper, the larger study will also include a combination PCC-CCC process using two of the processes being evaluated, a PCC process followed by FGD, and a CCC process followed by FGD. In the full study coal sulfur contents of 0.7, 2.0, and 3.5% will also be evaluated in addition to the 5.0% sulfur coal presented here. Premise conditions which differ from the FGD premises are (1) an update of the time base to a 1979-1982 construction period and a 1982 startup, (2) the use of a 2000-MW power plant for the base case, (3) a more detailed coal composition, and (4) a change in boiler operating time to 5500 hr/yr. Direct comparison to other evaluations should take these differences in consideration.

PROCESS DESCRIPTIONS

The three PCC processes represent widely used commercial technology and were selected for study because they offer a relatively high level of sulfur reduction compared to other PCC methods. The three CCC processes are not commercial processes, but have been developed to bench-scale or limited pilot-plant stages. Additional development could make significant changes in their technical, and thus economic, potential for sulfur reduction. The PCC processes are somewhat limited in their desulfurization application since they only remove pyritic sulfur. Two of the CCC processes remove significant quantities of organic sulfur in addition to pyritic sulfur.

Physical Coal Cleaning

PCC Process I. This process uses a dense-medium vessel for the coarse coal, a dense-medium cyclone for the intermediate-sized coal, and froth flotation for the fine coal, as shown in Figure 1.

The 3-inch x 0 coal is crushed and screened to three size fractions: 37% 2-inch x 3/8-inch coarse coal, 55% 3/8-inch x 28-mesh intermediate-sized coal, and 8% 28-mesh x 0 fine coal. The coarse coal is immersed

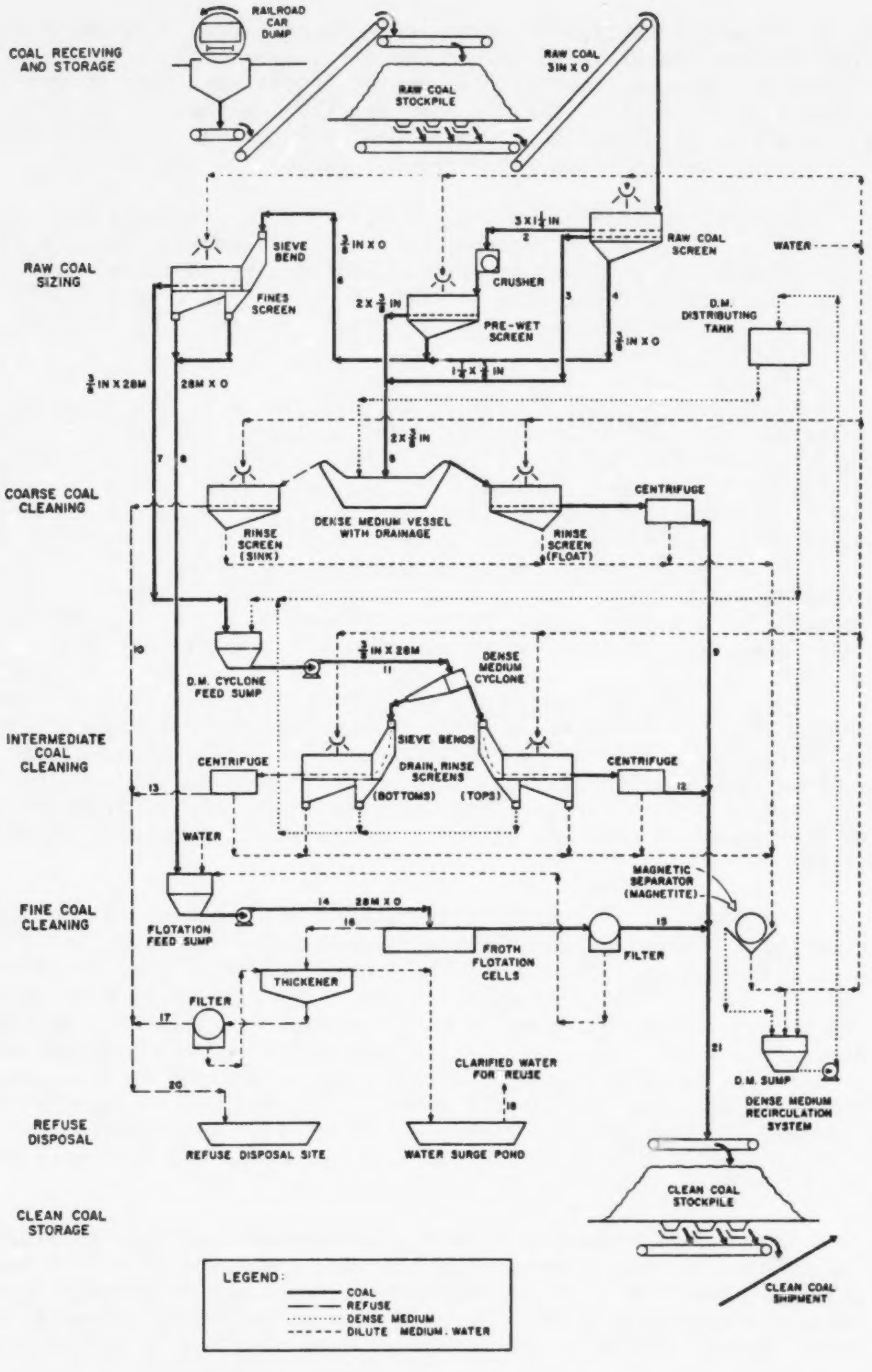

COAL RECEIVING AND STORAGE
RAILROAD CAR DUMP
RAW COAL STOCKPILE
RAW COAL 3IN X 0
RAW COAL SIZING
SIEVE BEND
$\frac{3}{8}$ IN X 0
FINES SCREEN
6
$3 \times 1\frac{1}{2}$ IN
2
RAW COAL SCREEN
WATER
CRUSHER
$2 \times \frac{3}{8}$ IN
PRE-WET SCREEN
3
4
$\frac{3}{8}$ IN X 0
D.M. DISTRIBUTING TANK
$1\frac{1}{2} \times \frac{3}{8}$ IN
$\frac{3}{8}$ IN X 28M
28M X 0
$2 \times \frac{3}{8}$ IN
7
8
5
COARSE COAL CLEANING
RINSE SCREEN (SINK)
DENSE MEDIUM VESSEL WITH DRAINAGE
RINSE SCREEN (FLOAT)
CENTRIFUGE
10
9
D.M. CYCLONE FEED SUMP
$\frac{3}{8}$ IN X 28M
11
DENSE MEDIUM CYCLONE
INTERMEDIATE COAL CLEANING
SIEVE BENDS
DRAIN, RINSE SCREENS
(BOTTOMS)
(TOPS)
CENTRIFUGE
CENTRIFUGE
13
12
WATER
FINE COAL CLEANING
MAGNETIC SEPARATOR (MAGNETITE)
14
28M X 0
16
15
FLOTATION FEED SUMP
FROTH FLOTATION CELLS
FILTER
THICKENER
FILTER
17
21
D.M. SUMP
DENSE MEDIUM RECIRCULATION SYSTEM
CLARIFIED WATER FOR REUSE
20
18
REFUSE DISPOSAL
REFUSE DISPOSAL SITE
WATER SURGE POND
CLEAN COAL STOCKPILE
CLEAN COAL STORAGE
CLEAN COAL SHIPMENT
LEGEND:
COAL
REFUSE
DENSE MEDIUM
DILUTE MEDIUM. WATER

Figure 1. Flow diagram for physical coal-cleaning process I.

in a 1.55 specific-gravity magnetite - water slurry in trough-type dense-medium vessels. The float fraction, about 85% of the feed, is rinsed and drained, dewatered in basket centrifuges, and conveyed to the clean coal stockpile. The 15% sink fraction is rinsed, drained, and discarded as refuse without centrifugal dewatering. Integral drain screens recover the magnetite from the clean coal and refuse.

The intermediate-sized coal is slurried in a pulping tank and cleaned in dense-medium cyclones at a nominal specific gravity of 1.55. The cyclones are followed by conventional drain-and-rinse screens and by basket centrifuge dewatering of both the clean coal float fraction, which is sent to the clean coal stockpile, and the sink fraction, which is discarded as refuse.

Single-stage froth flotation is used on the entire fine coal fraction. For flotation feed, a pulp density of 10% solids is formed in the flotation feed sump by diluting the coal slurry with filtrate from the flotation process clean coal filter. The float-fraction coal concentrate, at about 20% solids, is dewatered on rotary vacuum filters and conveyed to the clean coal stockpile. Froth flotation tailings flow to a thickener whose underflow is dewatered on a rotary vacuum filter and discarded as refuse. Filtrate is returned to the thickener for further settling to control slime. Thickener overflow flows to a clarified water pond from which water is returned for reuse in the plant.

PCC Process II. This process, shown in Figure 2, uses dense-medium cyclones operated at a relatively low specific gravity for the production of a limited overflow fraction of highly cleaned coal. The bottoms from the low-gravity cyclones are pumped to dense-medium cyclones operated at a high specific gravity for the production of "middling" (medium-quality) coal and refuse. The fine coal is recovered by froth flotation.

The raw coal is reduced by screening and crushing to two size fractions. The 3/4-inch x 28-mesh intermediate-sized coal constitutes 81% of the raw coal feed while the 28-mesh x 0 fine coal is 19% of the raw coal feed. The intermediate-sized coal is slurried with water and fed to the low-gravity dense-medium cyclones operated at a specific gravity of 1.34. About 49% of the intermediate-sized coal fed to the cyclones is taken off as an overflow. It is drained on sieve bends and vibrating screens, washed with water, again drained on the vibrating screens, centrifuged, and conveyed to the clean coal stockpile. Underflow from the cyclones is drained on sieve bends and vibrating screens for further processing.

The drained underflow from the low-gravity cyclones is slurried and fed to high-gravity dense-medium cyclones operating at a 1.55 specific gravity. Overflow from the high-gravity cyclones is drained on sieve bends and vibrating screens, washed with water on the vibrating screens, and dewatered in basket centrifuges. This overflow, a middling-quality

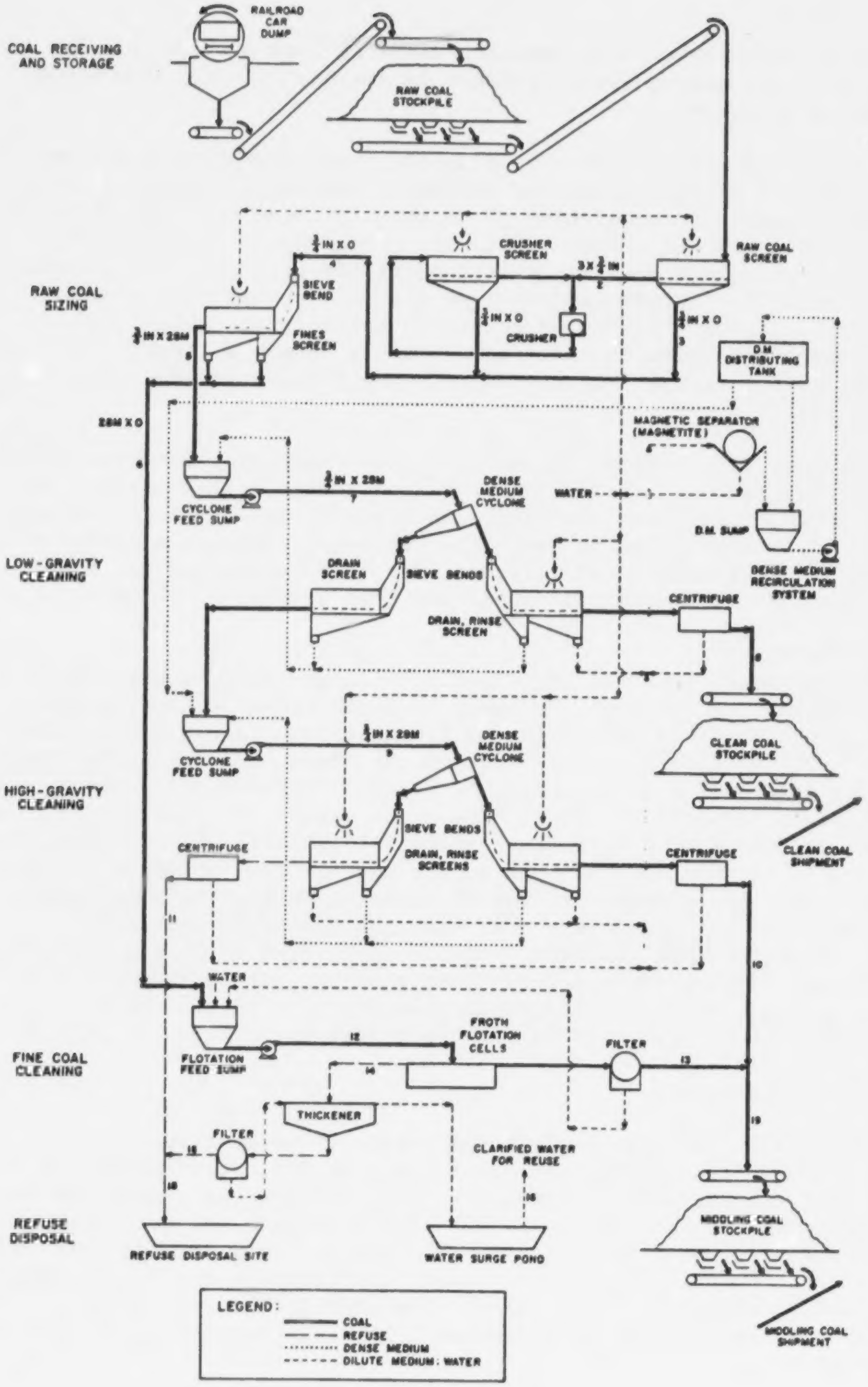

COAL RECEIVING AND STORAGE
RAILROAD CAR DUMP
RAW COAL STOCKPILE
RAW COAL SIZING
CRUSHER SCREEN
RAW COAL SCREEN
3/4 IN X 0
4
SIEVE BEND
FINES SCREEN
3/4 IN X 28M
5
3 X 3/4 IN
2
3/4 IN X 0
CRUSHER
3/4 IN X 0
3
D M DISTRIBUTING TANK
28M X 0
6
MAGNETIC SEPARATOR (MAGNETITE)
CYCLONE FEED SUMP
3/4 IN X 28M
7
DENSE MEDIUM CYCLONE
WATER
D.M. SUMP
DENSE MEDIUM RECIRCULATION SYSTEM
LOW-GRAVITY CLEANING
DRAIN SCREEN
SIEVE BENDS
DRAIN, RINSE SCREEN
CENTRIFUGE
8
CLEAN COAL STOCKPILE
CYCLONE FEED SUMP
3/4 IN X 28M
9
DENSE MEDIUM CYCLONE
HIGH-GRAVITY CLEANING
SIEVE BENDS
DRAIN, RINSE SCREENS
CENTRIFUGE
CENTRIFUGE
CLEAN COAL SHIPMENT
11
10
WATER
FROTH FLOTATION CELLS
FINE COAL CLEANING
FLOTATION FEED SUMP
12
FILTER
13
14
THICKENER
19
FILTER
15
CLARIFIED WATER FOR REUSE
18
16
REFUSE DISPOSAL
REFUSE DISPOSAL SITE
WATER SURGE POND
MIDDLING COAL STOCKPILE
MIDDLING COAL SHIPMENT
LEGEND:
COAL
REFUSE
DENSE MEDIUM
DILUTE MEDIUM: WATER

Figure 2. Flow diagram for physical coal-cleaning process II.

coal product, is conveyed to a separate middling coal stockpile. The underflow from the high-gravity cyclones is drained, rinsed, dewatered, and discarded as refuse.

The fine coal fraction is pulped in the froth flotation feed sump to a pulp density of 10% solids and pumped to the froth flotation cells. The flotation overflow, consisting of relatively clean coal, is filtered on rotary vacuum filters. Since this coal stream is of lower quality than the highly cleaned coal from the low-gravity cyclones, the flotation product is added to the middling coal stockpile with the coal from the high-gravity cyclones. Flotation underflow tailings are pumped to a thickener whose underflow is filtered on rotary vacuum filters. To control slimes the refuse filtrate is returned to the thickener for additional settling.

PCC Process III. This process uses dense-medium cyclones to clean 67% of the coal feed as a 1-1/2-inch x 8-mesh coarse coal fraction. A fine coal 8-mesh x 200-mesh fraction, amounting to 31% of the coal feed, is cleaned on concentrating tables. The remaining 200-mesh x 0 fine coal fraction is thickened and filtered without cleaning and added to the clean coal product. The flowsheet for this process is shown in Figure 3.

The coarse coal fraction is pulped and cleaned in 1.55 specific-gravity dense-medium cyclones, followed by conventional drain-and-rinse screening and mechanical dewatering of the underflow refuse product and the clean coal overflow, which is sent to the clean coal stockpile. The fine coal consists of a major stream of 8-mesh x 200-mesh size and a minor stream of 200-mesh x 0 size. The major stream is cleaned with concentrating tables at a water-to-coal ratio of 3:2 in the coal feed. Dressing water is also added along the top edge of the tables to provide stable flow across its deck. Including dressing water, the total water-to-coal ratio is 2:1. The clean coal from the tables is partially dewatered on sieve bends followed by final dewatering in basket centrifuges. The sieve bend filtrates are added, along with the dilute 200-mesh x 0 slurry, to a thickener. The thickener underflow is filtered on a rotary vacuum filter and added to the clean coal product.

Chemical Coal Cleaning

KVB Process. This process is the result of several years of research in chemical desulfurization of fuels by KVB, Incorporated, a Research-Cottrell Company. According to KVB, the process removes 90 to 99% of the pyritic sulfur and up to 40% of the organic sulfur in coal. The process was patented in September 1975 and has been demonstrated in bench-scale equipment. The process, shown in Figures 4 and 5, consists of a selective oxidation of the sulfur compounds in the coal using gaseous NO_2 in the presence of oxygen at a low temperature and atmospheric pressure.

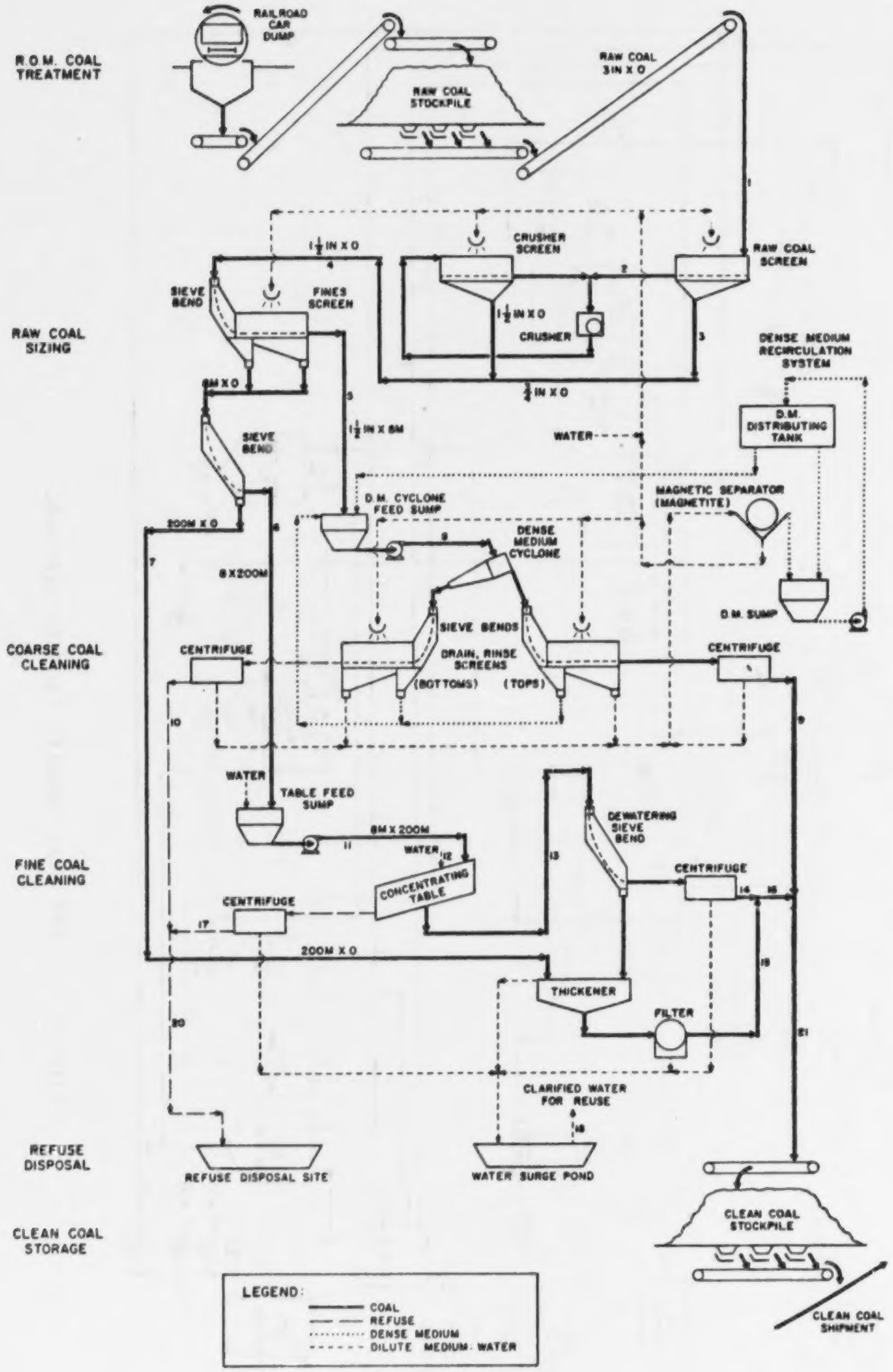

R.O.M. COAL TREATMENT
RAILROAD CAR DUMP
RAW COAL STOCKPILE
RAW COAL 3IN X O
1
RAW COAL SIZING
CRUSHER SCREEN
RAW COAL SCREEN
1½ IN X O
4
2
SIEVE BEND
FINES SCREEN
1½ IN X O
CRUSHER
3
DENSE MEDIUM RECIRCULATION SYSTEM
8M X O
½ IN X O
5
1½ IN X 8M
SIEVE BEND
WATER
D.M. DISTRIBUTING TANK
D.M. CYCLONE FEED SUMP
MAGNETIC SEPARATOR (MAGNETITE)
200M X O
6
8
DENSE MEDIUM CYCLONE
7
8 X 200M
D.M. SUMP
COARSE COAL CLEANING
CENTRIFUGE
SIEVE BENDS
DRAIN, RINSE SCREENS
(BOTTOMS)
(TOPS)
CENTRIFUGE
10
9
WATER
TABLE FEED SUMP
DEWATERING SIEVE BEND
8M X 200M
11
WATER
12
13
FINE COAL CLEANING
CONCENTRATING TABLE
CENTRIFUGE
CENTRIFUGE
14
16
17
200M X O
THICKENER
15
FILTER
20
21
CLARIFIED WATER FOR REUSE
18
REFUSE DISPOSAL
REFUSE DISPOSAL SITE
WATER SURGE POND
CLEAN COAL STORAGE
CLEAN COAL STOCKPILE
LEGEND:
COAL
REFUSE
DENSE MEDIUM
DILUTE MEDIUM WATER
CLEAN COAL SHIPMENT

Figure 3. Flow diagram for physical coal-cleaning process III.

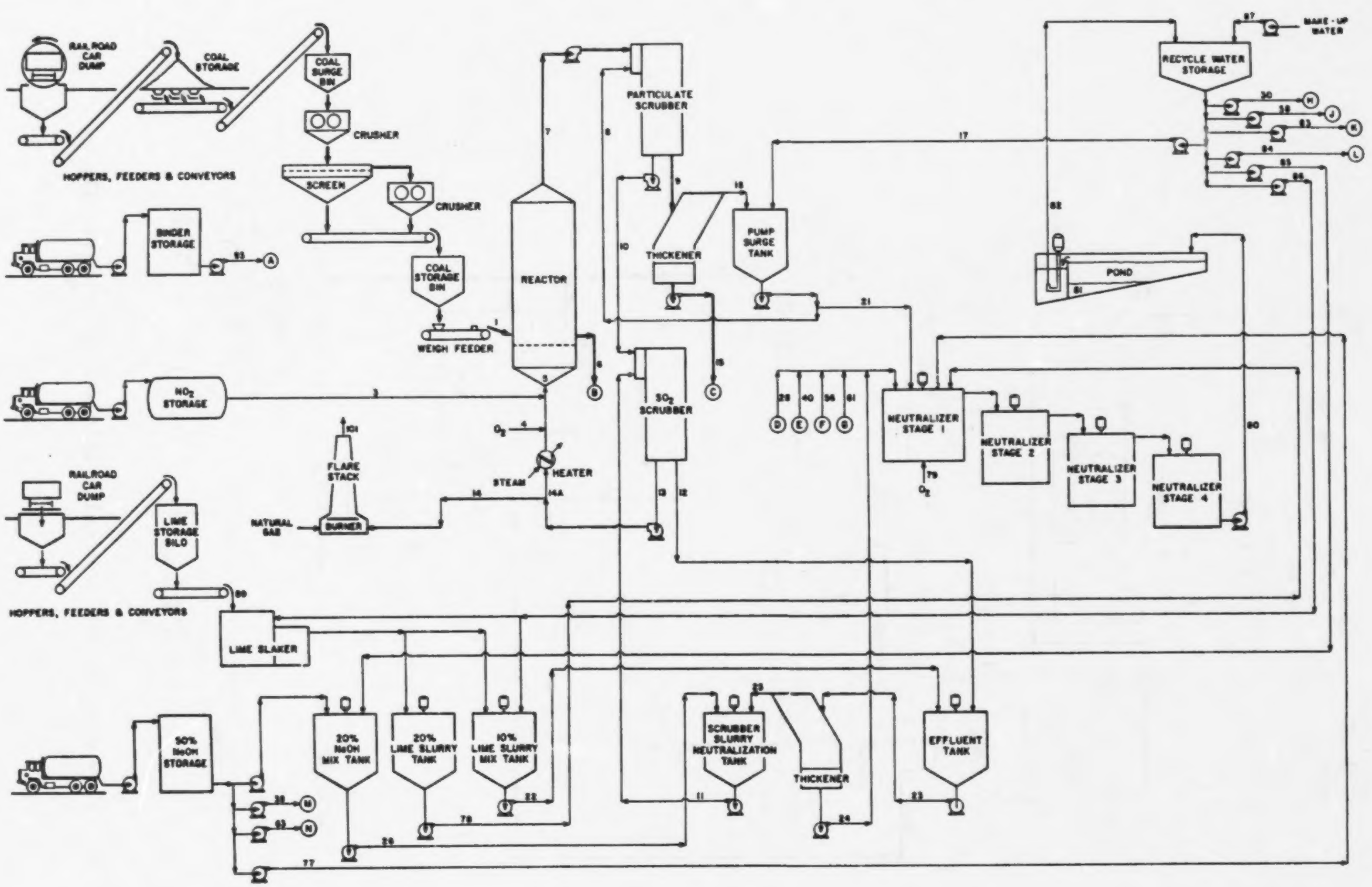

RAILROAD CAR DUMP
COAL STORAGE
HOPPERS, FEEDERS & CONVEYORS
COAL SURGE BIN
CRUSHER
SCREEN
CRUSHER
COAL STORAGE BIN
WEIGH FEEDER
BINDER STORAGE
A
NO_2 STORAGE
REACTOR
B
O_2
HEATER
STEAM
FLARE STACK
NATURAL GAS
BURNER
RAILROAD CAR DUMP
LIME STORAGE SILO
HOPPERS, FEEDERS & CONVEYORS
LIME SLAKER
50% NaOH STORAGE
M
N
20% NaOH MIX TANK
20% LIME SLURRY TANK
10% LIME SLURRY MIX TANK
PARTICULATE SCRUBBER
THICKENER
PUMP SURGE TANK
C
SO_2 SCRUBBER
D
E
F
G
NEUTRALIZER STAGE 1
O_2
NEUTRALIZER STAGE 2
NEUTRALIZER STAGE 3
NEUTRALIZER STAGE 4
SCRUBBER SLURRY NEUTRALIZATION TANK
THICKENER
EFFLUENT TANK
POND
RECYCLE WATER STORAGE
MAKE-UP WATER
H
J
K
L

Figure 4 . KVB coal desulfurization process.

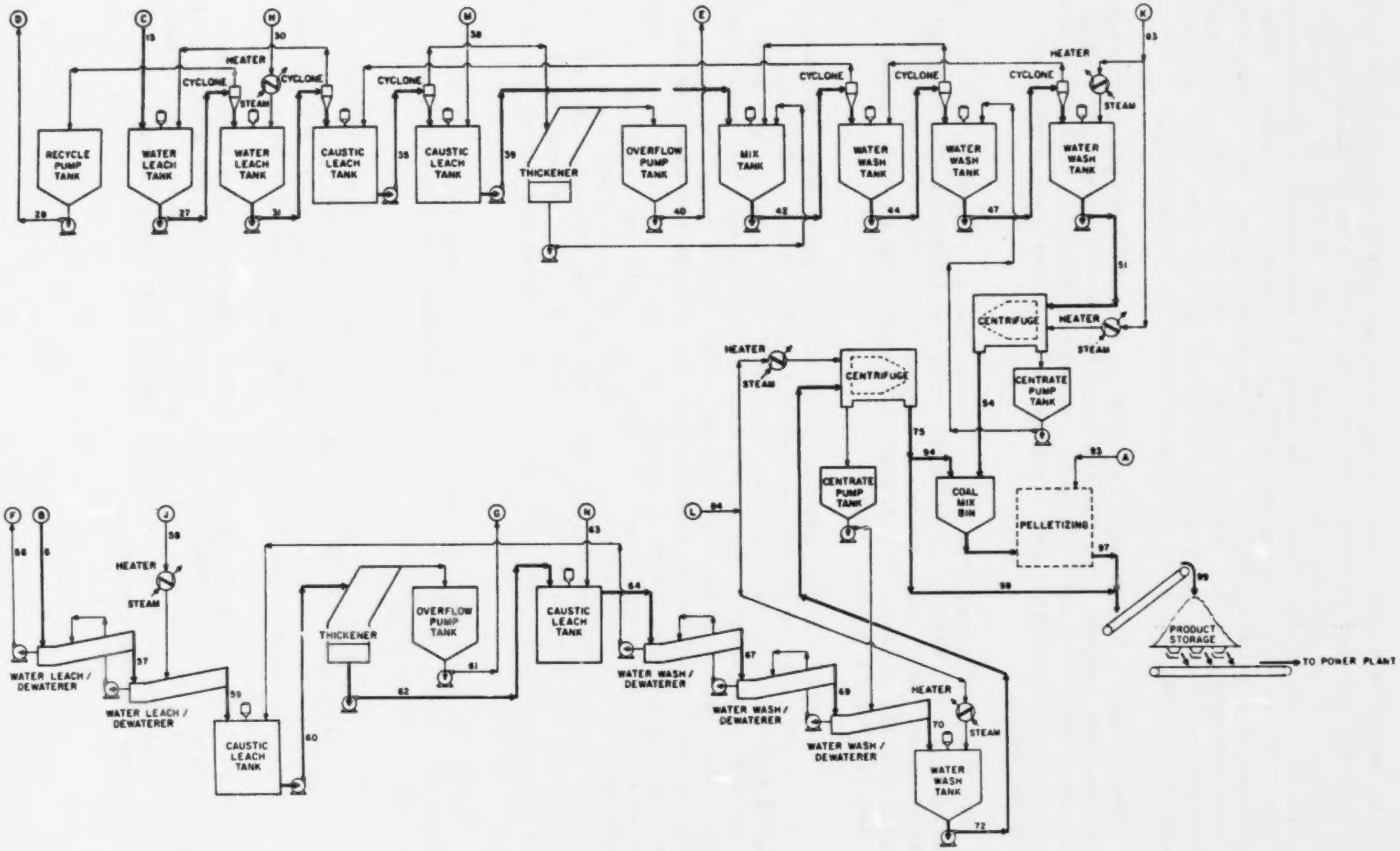

D
C
15
H
30
M
38
E
K
83
HEATER
CYCLONE
STEAM
RECYCLE PUMP TANK
WATER LEACH TANK
WATER LEACH TANK
CAUSTIC LEACH TANK
CAUSTIC LEACH TANK
THICKENER
OVERFLOW PUMP TANK
MIX TANK
WATER WASH TANK
WATER WASH TANK
WATER WASH TANK
29
27
31
35
39
40
42
44
47
51
CENTRIFUGE
CENTRATE PUMP TANK
COAL MIX BIN
PELLETIZING
54
75
94
93
A
97
98
99
PRODUCT STORAGE
TO POWER PLANT
L
84
F
B
J
56
6
58
37
59
60
61
62
G
N
63
64
67
69
70
72
WATER LEACH / DEWATERER
WATER LEACH / DEWATERER
CAUSTIC LEACH TANK
THICKENER
OVERFLOW PUMP TANK
CAUSTIC LEACH TANK
WATER WASH / DEWATERER
WATER WASH / DEWATERER
WATER WASH / DEWATERER
WATER WASH TANK

Figure 5. KVB coal desulfurization process (continued).

The 3-inch x 0 coal from the raw coal stockpile is crushed to 1/4-inch x 0 size and fed to fluidized-bed reactors. Hot oxidizing gas, containing 5% NO_2, 2.5% O_2, N_2, H_2O, and a trace of SO_2, is circulated through the coal in the reactor and oxidizes the sulfur to sulfates and SO_2 gas. The reactions occur at 200°F and atmospheric pressure. The reactions are exothermic but do not provide sufficient heat to maintain the reaction temperature. The oxidation is carried out at a low oxygen concentration so that the reaction effluent gas is very low in NO_2 and O_2 and high in NO, which is reduced to N_2 in the flare stack.

The 28-mesh x 0 fine coal fraction is entrained by the oxidizing gas stream and removed from the reactor off-gas stream by particulate scrubbers. The 1/4-inch x 28-mesh coarse coal fraction is removed from the bottom of the reactor and transferred to coarse coal washing and leaching trains.

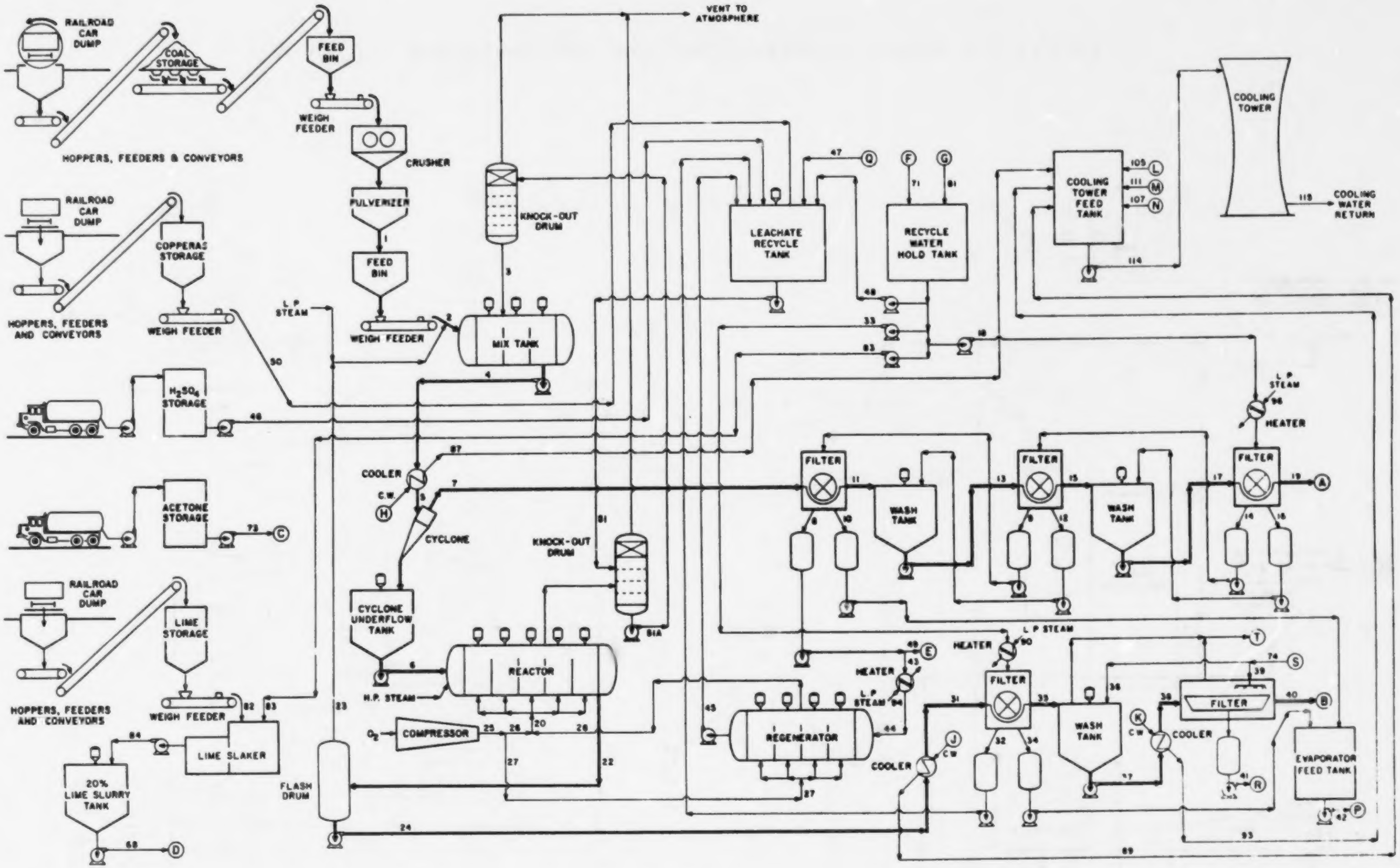
The SO_2 is removed from the oxidizing gas stream in venturi scrubbers by scrubbing with Na_2SO_3. The $NaHSO_3$ solution formed in the scrubbers is then treated with slaked lime to regenerate Na_2SO_3 and produce a calcium sulfite sludge. The sludge is further treated with oxygen in the neutralizer to produce gypsum.

The fine coal slurry from the particulate scrubbers is increased to 37% solids in a thickener and leached with 200°F water. The water-leached fine coal is then leached with 200°F NaOH solution and again washed with 200°F water. Cyclone classifiers, followed by centrifuges, dewater the coal to about 10% moisture. The coarse coal from the fluidized-bed reactors is processed by the same method used for the fine coal washing and leaching except that spiral classifiers are used instead of tanks and cyclones. The hot water wash and leaching solutions are treated in the neutralizer with slaked lime to produce a waste sludge of gypsum and sodium jarosite. All of the fine coal and 15% of the coarse coal is pelletized. The pelletized coal, containing 5% moisture, is combined with the unpelletized portion and stored in open-air stockpiles.

Potential problems with the KVB process are possible explosion hazards involved in dry oxidation of pulverized coal and possible nitrogen uptake in the coal.

TRW "Gravichem" Process. The "Gravichem" coal desulfurization process was developed by TRW, who claim the process will remove 90 to 95% of the pyritic sulfur but none of the organic sulfur. The process has been demonstrated in an 8 ton/day plant at a TRW test site.

The process, shown in Figures 6 and 7, consists of a sink-float gravity separation, followed by selective oxidation of the sink fraction with $Fe_2(SO_4)_3$, followed by acetone leaching. The leaching and regeneration reactions are both exothermic but do not supply enough heat to maintain the reactor temperature.


RAILROAD CAR DUMP
COAL STORAGE
HOPPERS, FEEDERS & CONVEYORS
FEED BIN
WEIGH FEEDER
CRUSHER
PULVERIZER
1
FEED BIN
WEIGH FEEDER
2
VENT TO ATMOSPHERE
KNOCK-OUT DRUM
3
MIX TANK
4
RAILROAD CAR DUMP
COPPERAS STORAGE
HOPPERS, FEEDERS AND CONVEYORS
WEIGH FEEDER
L.P. STEAM
50
H2SO4 STORAGE
46
ACETONE STORAGE
73
C
LEACHATE RECYCLE TANK
47
Q
F
G
71
81
RECYCLE WATER HOLD TANK
48
33
83
18
COOLING TOWER FEED TANK
105
L
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N
114
COOLING TOWER
115
COOLING WATER RETURN
COOLER
C.W.
H
87
5
7
CYCLONE
CYCLONE UNDERFLOW TANK
6
H.P. STEAM
KNOCK-OUT DRUM
91
91A
REACTOR
O2
COMPRESSOR
25
26
20
28
27
22
23
FLASH DRUM
24
RAILROAD CAR DUMP
LIME STORAGE
HOPPERS, FEEDERS AND CONVEYORS
WEIGH FEEDER
82
83
LIME SLAKER
84
20% LIME SLURRY TANK
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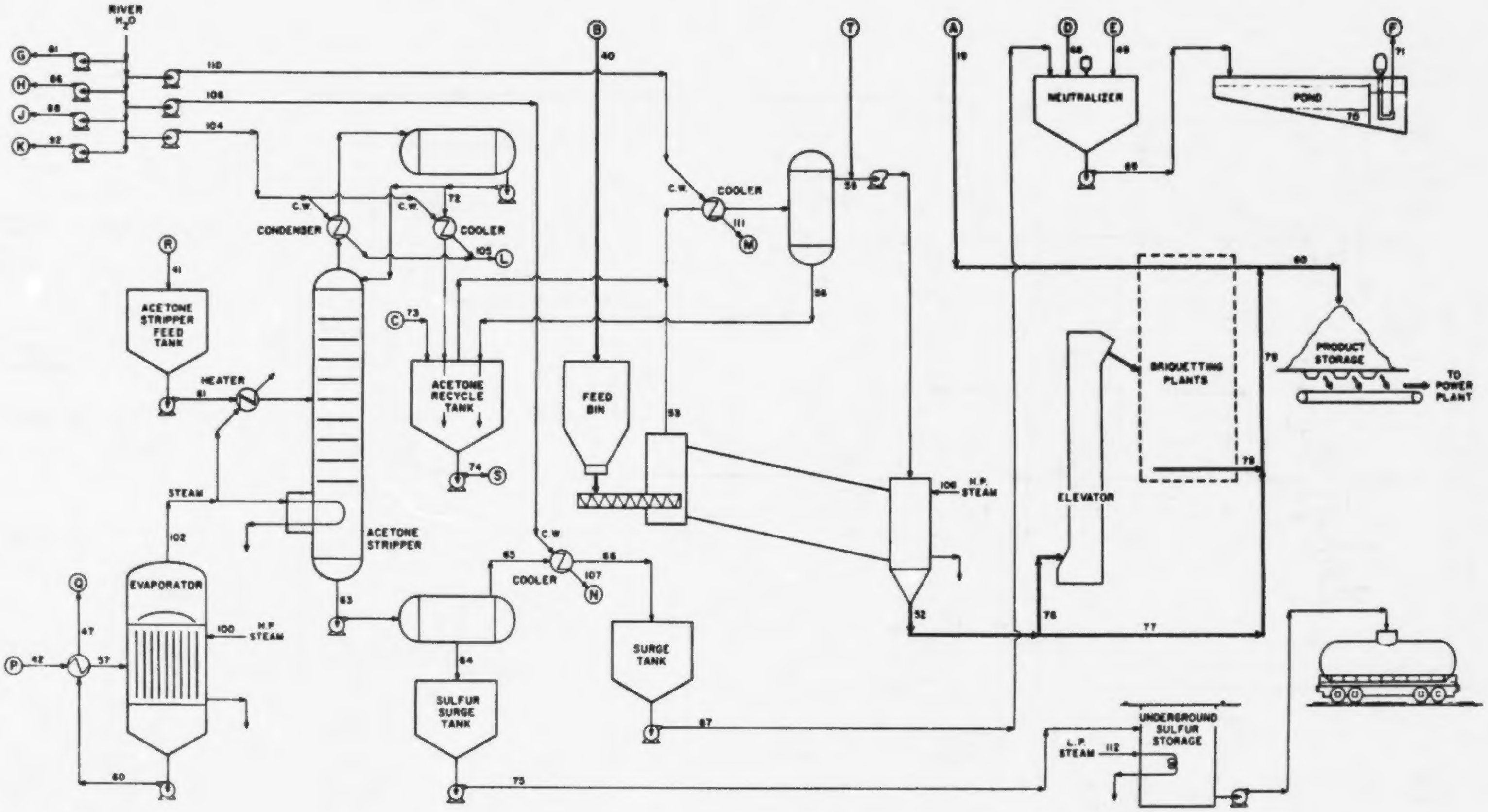
Figure 6 . TRW-"Gravichem" coal desulfurization process.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E
69
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71
NEUTRALIZER
POND
70
89
C.W.
CONDENSER
C.W.
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COOLER
105
L
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41
ACETONE
STRIPPER
FEED
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HEATER
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ACETONE
RECYCLE
TANK
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FEED
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C.W.
COOLER
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56
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108
H.P.
STEAM
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BRIQUETTING
PLANTS
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PRODUCT
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TANK
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C.W
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66
COOLER
107
N
SURGE
TANK
67
UNDERGROUND
SULFUR
STORAGE
L.P.
STEAM
112

Figure 7. TRW-"Gravichem" coal desulfurization process (continued).

The raw coal is crushed to 14-mesh top size and the crushed coal is slurried in recycled leach solution containing 7.5% total iron as $FeSO_4$ and $Fe_2(SO_4)_3$, plus 4% H_2SO_4. The slurry is cooled to control the specific gravity at 1.31 and pumped to cyclones for a sink-float separation. The cyclone float fraction, which contains about 32% of the total coal and has a low pyritic concentration, is filtered, washed, and conveyed to the clean coal stockpile.

The sink fraction is pumped to reactors operating at 250°F and 35 psig with a 6-hour residence time. The oxidation reaction produces $FeSO_4$, H_2SO_4, and sulfur. The $Fe_2(SO_4)_3$ solution is regenerated by sparging with oxygen. The reacted coal slurry is cooled, filtered, and washed with water. The filtered coal is then slurried with acetone, cooled to 85°F, and filtered. The acetone leaching removes most, but not all, of the sulfur in the coal, which is recovered from the stripper bottoms as a marketable byproduct. The coal is dried and the acetone recovered for recycling. Approximately 80% of this coal is hot briquetted, combined with the remainder of the dried coal and the float coal product, and conveyed to the clean coal stockpile.

The strong leachate bleedstream from the filter and the bottoms from the stripper are neutralized with slaked lime and the neutralized slurry pumped to a settling pond.

Potential problems with the TRW process include the presence of a very corrosive dilute sulfuric acid and iron sulfate solution and potential environmental problems associated with the disposal of the gypsum - iron hydroxide sludge.

Kennecott Process. Kennecott Copper Corporation began development of this process in 1970. Development continued through May 1975, during which time the process was demonstrated at a bench-scale level. The process, shown in Figure 8, consists of an oxidation system in which a portion of the sulfur in the coal is oxidized to soluble sulfates by sparging oxygen through pulverized coal at a high temperature and pressure. The reaction is exothermic and provides sufficient heat to maintain the reaction temperature. The soluble sulfates are removed by washing and neutralized with slaked lime to produce a waste sludge of $Fe(OH)_2$ and gypsum.

The coal is pulverized to 80% 100 mesh x 0 with crushers and wet ball mills and the slurry is heated and pumped to agitated reactors. The reactors operate at 350°F and 315 psig with a 1-hour hold time. The reacted slurry is cooled, thickened to 35% solids, and water washed on rotary drum filters.

Eighty percent of the washed coal is pelletized. The pelletized coal is then combined with the unpelletized portion and conveyed to the clean coal stockpile.

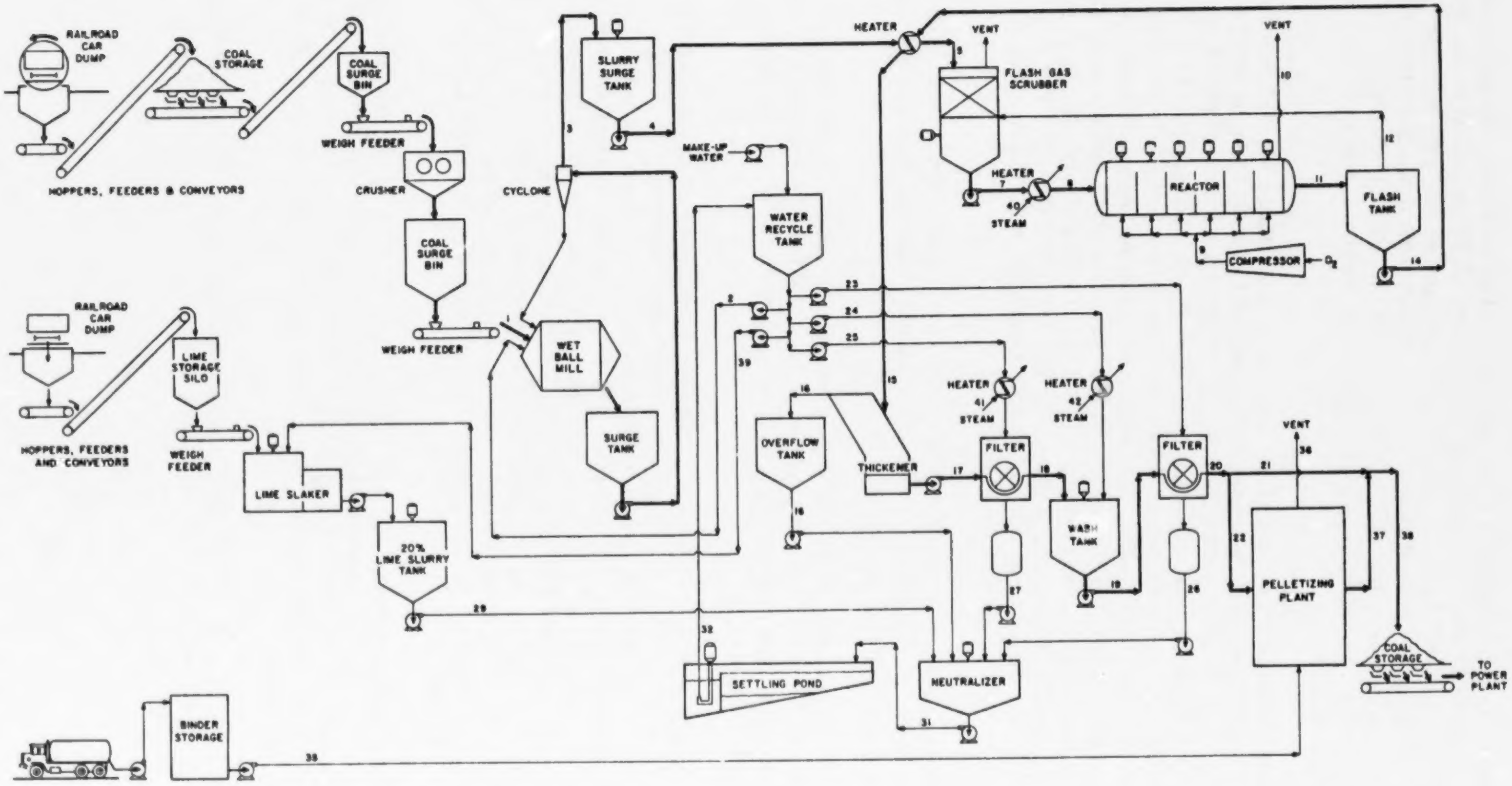
RAILROAD CAR DUMP
COAL STORAGE
COAL SURGE BIN
WEIGH FEEDER
HOPPERS, FEEDERS & CONVEYORS
CRUSHER
COAL SURGE BIN
WEIGH FEEDER
CYCLONE
SLURRY SURGE TANK
WET BALL MILL
SURGE TANK
MAKE-UP WATER
WATER RECYCLE TANK
HEATER
VENT
FLASH GAS SCRUBBER
HEATER
STEAM
REACTOR
FLASH TANK
COMPRESSOR
O_2
VENT
RAILROAD CAR DUMP
HOPPERS, FEEDERS AND CONVEYORS
LIME STORAGE SILO
WEIGH FEEDER
LIME SLAKER
20% LIME SLURRY TANK
OVERFLOW TANK
THICKENER
HEATER
STEAM
FILTER
WASH TANK
HEATER
STEAM
FILTER
VENT
PELLETIZING PLANT
COAL STORAGE
TO POWER PLANT
SETTLING POND
NEUTRALIZER
BINDER STORAGE
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Figure 8. Kennecott coal desulfurization process.

The clear liquid from the thickeners and filters, containing $FeSO_4$ and H_2SO_4, is pumped to a neutralizer where it is treated with slaked lime. The neutralized slurry of gypsum and iron hydroxide is pumped to a settling pond from which supernate water is returned for use in the process.

Potential problems with the Kennecott process include the presence of a very corrosive dilute H_2SO_4-$FeSO_4$ solution, reactor design limitations due to high operating pressures, and potential environmental problems associated with the disposal of the gypsum - iron hydroxide sludge.

RESULTS OF PHYSICAL AND CHEMICAL COAL-CLEANING STUDY

Cleaning Performance

The cleaning performances of the six coal cleaning processes at the base-case operating conditions described in the premises are shown in Table 1. The PCC processes are limited to removal of pyritic sulfur and have a considerably lower sulfur removal efficiency than the CCC processes. In addition, there is a significant weight reduction from removal of noncoal minerals (as well as some coal) in the PCC processes. The CCC processes remove most of the pyritic sulfur and the KVB and Kennecott processes remove up to 30 to 40% of organic sulfur.

TABLE 1. CLEANING PERFORMANCE OF PHYSICAL AND CHEMICAL

COAL-CLEANING PROCESSES - 5% SULFUR COAL

(Moisture-free basis)

		Physical coal cleaning			Chemical coal cleaning		
	Raw coal	PCC-I	PCC-II	PCC-III	KVB	TRW Gravichem	Kennecott
Total sulfur, %	5.00	3.67	3.51	3.78	1.30	1.86	1.80
Pyritic sulfur, %	3.35	2.02	1.86	2.13	0.09	0.09	0.40
Ash, %	16.7	10.1	9.3	10.6	11.4	13.6	15.6
Btu/lb	12,000	13,000	13,100	12,900	12,600	12,300	12,000
Btu recovery, %	-	90.7	91.4	90.7	99.4	97.7	94.3
Weight recovery, %	-	84.2	84.0	84.7	94.7	95.3	102.9
Total sulfur, lb/MBtu	4.17	2.84	2.68	2.93	1.00	1.50	1.50
Sulfur removed, % Btu basis	-	32	36	30	76	64	64

Sulfur removal efficiencies for the physical processes range from 30 to 36% on the basis of raw and cleaned coal heating values. Weight and Btu recoveries are about 84% and 91% respectively. There is also an increase in cleaned coal heating value and a reduction in ash. Removal efficiencies of the chemical processes range from 65 to 79% with no appreciable weight or Btu loss. There is also less increase in heating value and less reduction in ash content compared to the physical processes.

In comparing the economics of these coal-cleaning processes the removal efficiencies must be considered. In contrast, FGD system economics are based on sulfur removal to a constant level and removal efficiencies do not enter as a cost variable.

Base-Case Economics

Base-case capital investment and annual revenue requirement breakdowns for the six processes are shown in Appendix B. A summary of these data and the removal efficiencies is shown in Table 2. Costs of a limestone scrubbing FGD system with pond disposal of sludge are also included in Table 2 for comparison with the coal-cleaning processes. All of the cost data are based on processes serving a 2000-MW power plant burning coal with 5% sulfur, as described in the premises.

The economic results and conclusions presented here are preliminary, based on data obtained thus far in the continuing coal-cleaning evaluation. Further data on combination processes and case variations will better amplify and define the economics of these processes, perhaps modifying some details of these results.

TABLE 2. PHYSICAL AND CHEMICAL COAL-CLEANING ECONOMIC DATA SUMMARY

Process	% sulfur reduction	Capital investment		Annual revenue requirements	
		$/kW	¢/lb sulfur removed/yr	Mills/kWh	¢/lb sulfur removed
PCC-I	32	34	36	2.7	16
PCC-II	36	40	40	2.9	16
PCC-III	30	39	45	2.9	18
KVB	79	86	49	9.2	26
TRW	65	114	77	7.4	27
Kennecott	65	141	85	14.7	49
FGD	85	119	68	5.6	18

Basis

2,000 MW, 5.0% sulfur in coal, 5,500 hr/yr, 9,500 Btu/kWh heat rate. FGD is limestone scrubbing, 25% scrubber redundancy, with pond sludge disposal. Percent sulfur reduction based on raw and cleaned coal heating values.

The three physical processes have capital investments of 34 to 40 \$/kW and annual revenue requirements of 2.7 to 2.9 mills/kWh. Considering removal efficiencies the capital investments are 36 to 45 ¢/lb S removed/yr and the annual revenue requirements are 16.3 to 18.3 ¢/lb S removed. In direct investment the major cost differences are in lower coal-cleaning equipment costs for the PCC I process, higher coal storage costs for the PCC II process, and slightly higher coal sizing costs for the PCC III process. Refuse disposal and land costs for the three processes did not differ greatly.

Annual revenue requirements for the physical processes do not differ greatly. Large direct costs such as coal loss, labor, and maintenance are similar for all three processes. Higher utility costs, particularly electricity, account for slightly higher total direct costs for the PCC II process. Lower capital charges account for most of the lower annual revenue requirements of the PCC I process. When sulfur removal efficiencies are also considered, however, the position of the PCC II process is improved while the PCC III process becomes the most expensive process.

The three chemical processes all have larger capital investments and annual revenue requirements than the physical processes and also differ more widely among themselves. Capital investments for the chemical processes range from 86 to 141 \$/kW and annual revenue requirements range from 9.2 to 14.7 mills/kWh. Considering removal efficiencies the range in costs is increased. Capital investments range from 49 to 85 ¢/lb S removed/yr and annual revenue requirements range from 26 to 49 ¢/lb S removed, with KVB process having both the lowest capital investment and annual revenue requirements.

In capital investment there are large differences in equipment costs for the chemical processes. The reactor - regenerator and acetone leaching and recovery costs are large elements in the TRW process. The reactor area, filtration, and agglomerization costs are large elements in the Kennecott process. The KVB process with an atmospheric reaction at low temperature has no similar high-cost areas and consequently has the lowest capital investment. When removal efficiencies are also considered, the position of the KVB process is further enhanced by its high efficiency, making it only slightly higher in capital investment, on the basis of cost versus sulfur removed, than the physical processes.

All of the chemical processes have much higher annual revenue requirements than the physical processes, primarily because of large conversion costs, indirect costs related to capital investment, and, in the KVB and Kennecott processes, large raw material costs. The relatively low annual revenue requirements of the TRW process are largely a result of low raw material costs. The KVB process, although it has the lowest conversion costs, has much higher raw material costs--more than half for NaOH. The Kennecott process combines high raw material costs with high conversion costs, particularly for steam and electricity, to produce the highest annual revenue requirements of the processes evaluated. When removal efficiencies are considered, differences between the KVB process and the TRW process are reduced. In all cases, however, the chemical processes remain more expensive to operate than the physical processes.

For comparison, a limestone scrubbing FGD unit with 85% sulfur reduction is included in Table 2. The capital investment for the FGD unit is much higher than those for the PCC processes and higher than the capital investment of the KVB process. Annual revenue requirements for the FGD system are higher than those of the physical processes although lower than those of the chemical processes.

When compared on the basis of cost in terms of sulfur removed, however, the capital investment of the FGD system is greatly reduced relative to the coal-cleaning processes. On this basis the KVB process with its high removal efficiency and relatively low capital investment compares favorably with the FGD system in capital investment. The PCC processes also remain less costly in capital investment than the FGD system in terms of cost versus sulfur removed. The annual revenue requirements of the FGD system are more than those of the PCC processes in terms of cost versus sulfur removed but remains lower than the annual revenue requirements of all the chemical processes.

Effect of Coal Sulfur Content on Economics

Figure 9 shows the effect of coal sulfur content on the capital costs of the six coal-cleaning processes and FGD. Annual revenue requirements are shown in Figure 10. Since the processes remove different percentages of sulfur from the feed coal, capital and operating costs per kilowatthour are not comparable on a direct basis. The cost comparisons are shown on the basis of quantity of sulfur removed, which incorporates removal efficiency. The three PCC processes and the KVB process have lower capital costs per pound of sulfur removed per year than FGD. The TRW and Kennecott processes have higher capital costs per pound of sulfur removed per year than FGD.

The three PCC processes have annual revenue requirements per pound of sulfur removed similar to FGD except for the 0.7% coal. All of the CCC processes have higher annual revenue requirements per pound of sulfur removed than FGD, with the Kennecott process being the highest.

Other Economic Benefits and Penalties of Using Cleaned Coal

In evaluating the capital investment and annual revenue requirements associated with coal cleaning, it is useful to also assess the other economic benefits and penalties for users of cleaned coal. In addition to the primary benefit that the cleaned coal is lower in pyritic and, depending on the process, organic sulfur, it is generally also lower in ash and higher in calorific value, although often higher in surface moisture. Combustion of coal with these characteristics has numerous benefits as well as certain disadvantages to the user.

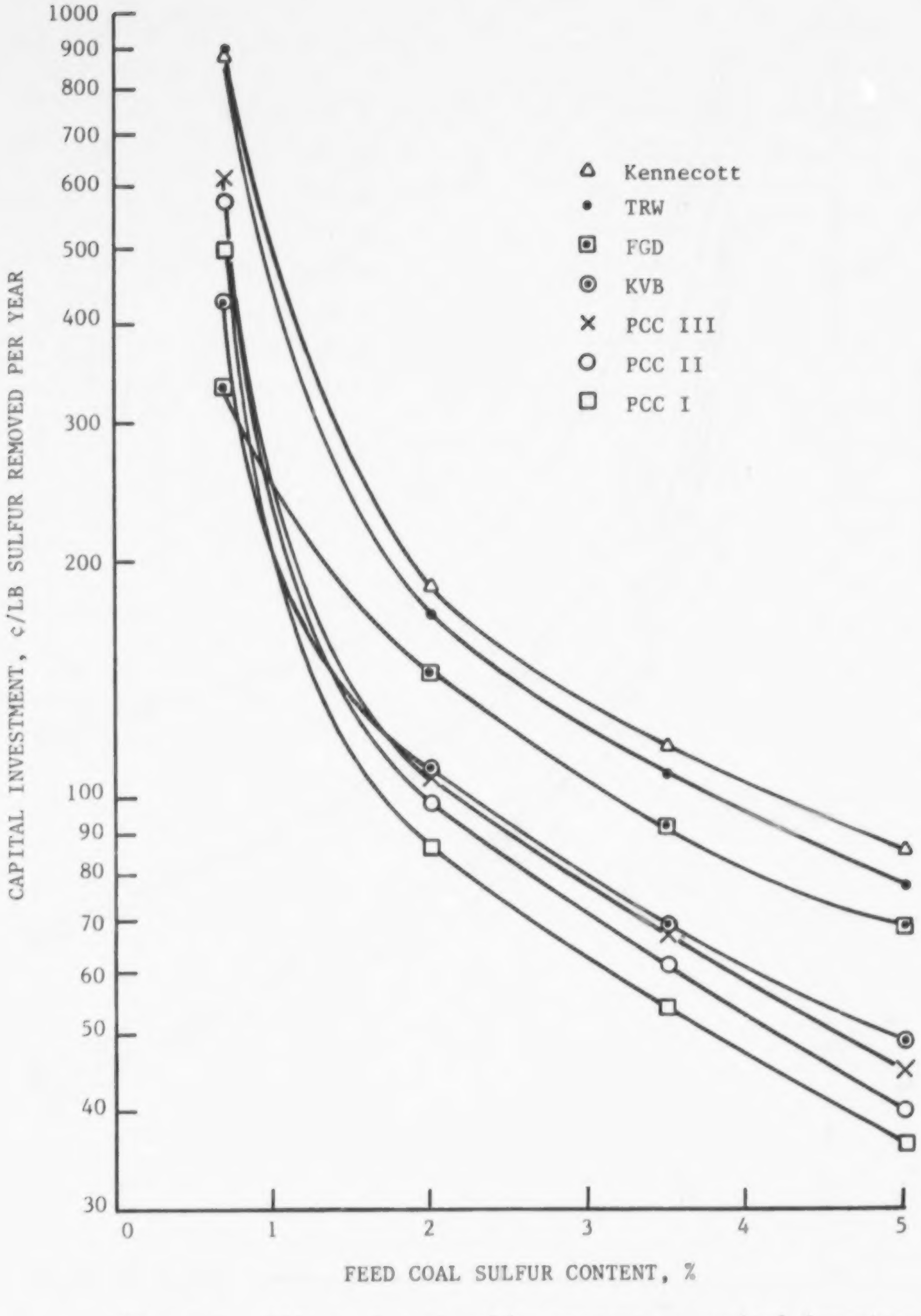

Figure 9. Effect of coal sulfur content on capital investment.

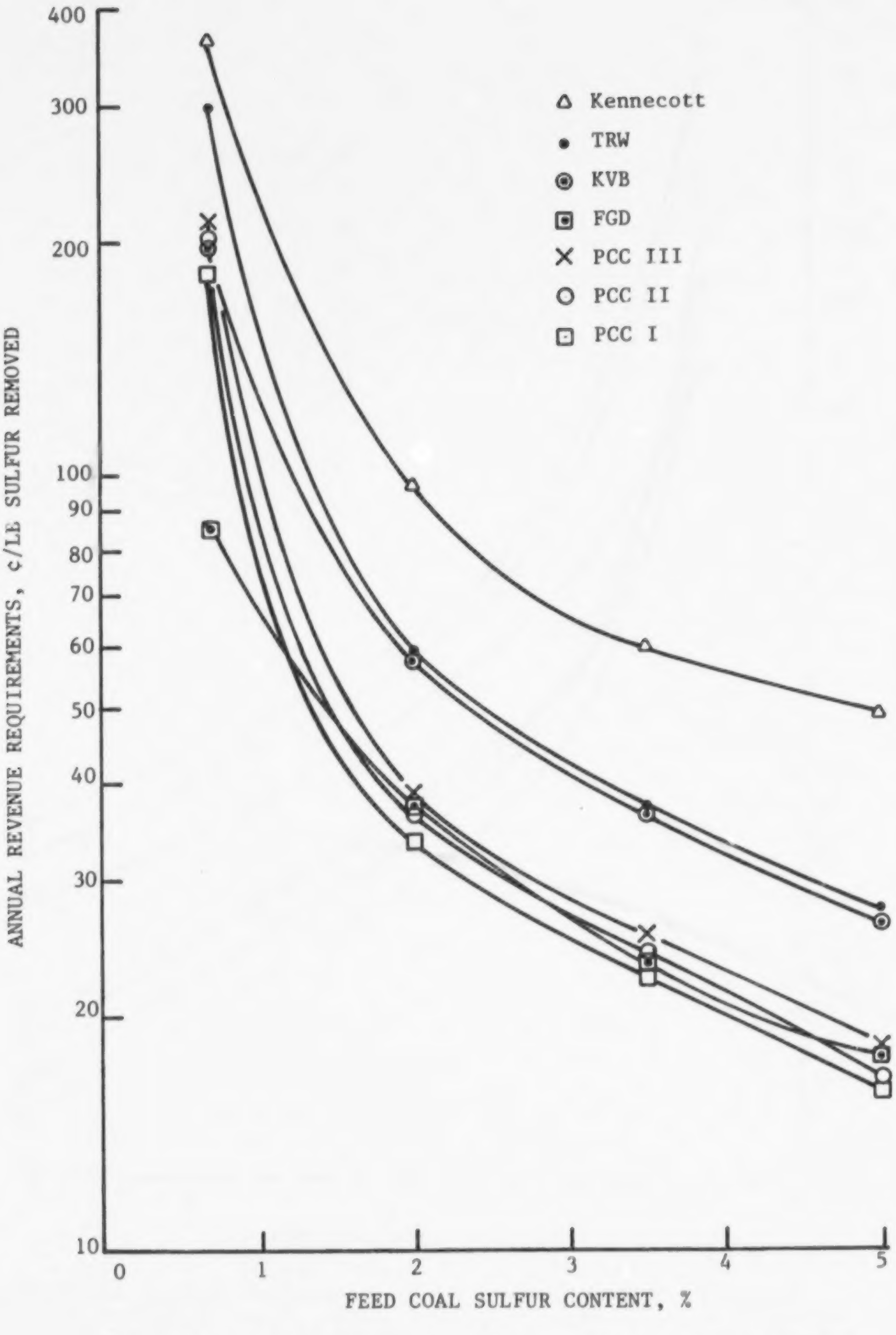

Figure 10. Effect of coal sulfur content on annual revenue requirements.

The net effect, however, is a credit which may be of sufficient magnitude to offset some of the increased cost of cleaned coal. Several of the significant economic effects of using cleaned coal are discussed below.

Transportation Costs. Coal beneficiation, if at the mine, decreases the cost of coal transportation by increasing the calorific value of the coal, consequently reducing the quantity of coal necessary to supply a given heat requirement.

Pension and Benefit Trust Fund. Provisions of the 1978 UMW contract require payment by the mine operator of $1.385 to the UMW Pension and Benefit Trust Fund for each ton of coal shipped to a consumer. If the coal-cleaning plant is at the mine, the cleaned coal will be higher in calorific value requiring a smaller tonnage to supply a required heat requirement, thus reducing this payment.

Pulverization Costs. PCC, by reducing mineral matter, decreases coal hardness and facilitates crushing. The increased calorific value of clean coal also reduces the quantity of coal to be crushed. The size of the clean coal product is considerably smaller than that of raw coal so that significant pulverization costs, which are already covered in the coal-cleaning costs, are saved. Detrimentally, cleaning may contribute additional surface moisture which makes pulverization more difficult.

Boiler Capacity. The higher calorific value of cleaned coal decreases the possibility that the utility boiler capacity will be derated because of deteriorating coal quality. Also, by reducing the slagging tendency of the coal, coal cleaning can permit the design of furnaces with higher heat transfer rates and correspondingly smaller furnace volume.

Boiler Performance. Cleaned coal can improve boiler performance by reducing slagging, fouling, and corrosion problems. This can significantly reduce the cost of boiler operation and maintenance and increase the availability of the generating facility.

Ash Handling. Ash handling and disposal costs are decreased since coal cleaning generally reduces the total amount of ash handled. Less sensible heat is lost in the bottom ash because of the lower ash levels.

FGD Operation. FGD systems generally have markedly better operation with low-sulfur coals. When coal cleaning is followed by FGD scrubbing, the lower sulfur level of the cleaned coal should give less FGD system downtime and a better overall utility availability.

FGD Capital and Operating Costs. FGD systems for boilers burning high-sulfur coal have higher capital costs because of the necessity for a large absorbent preparation facility, scrubber system, and area for sludge disposal. Corresponding savings in operating costs should also be realized with the low-sulfur cleaned coal.

ESP Size and Cost. The resistivity of fly ash is a major factor in determining the collection area of the ESP. Resistivity is determined by many factors, including ash composition and SO_3 level in the gas. With conventional ESP units, the removal of fly ash will generally be more difficult with low sulfur levels of cleaned coal and ESP costs may be increased. Other systems such as hot side ESP, bag filters, or pulsed ESP may be less expensive in certain cases when burning clean coal.

Surface Moisture. Higher moisture levels in cleaned coal resulting from the smaller particle sizes increase transportation costs and result in a heat loss when the water is heated and vaporized during the combustion process.

It is obvious that additional work is needed to quantify the economic magnitude of these benefits and penalties to the utility that uses cleaned coal.

Energy Requirements

Energy usages for the six processes are shown in Table 3. The comparison is made on the basis of total energy input consisting of raw coal feed and utilities. In addition to the electrical, steam, diesel fuel, or natural gas energy consumed in the process, there are other energy losses or usages that are specific for certain systems. Since the product coals are generally of finer size than the raw coal feed, they will have a higher surface moisture. Additional energy is needed to vaporize this extra moisture and to heat the water vapor to stack temperature. The three PCC processes have a significant Btu loss because part of the coal is entrapped in the refuse stream and discarded.

TABLE 3. PHYSICAL AND CHEMICAL COAL-CLEANING

ENERGY USAGE AND LOSSES

	PCC-I	PCC-II	PCC-III	KVB	TRW	Kennecott
Total energy input, 10^{12} Btu/yr	115.6	115.2	115.3	110.5	114.2	125.6
Energy Lost or Used						
Coal lost or used, % of input	9.3	8.7	9.2	0	0	1.2
Moisture increase in product coal	0.2	0.5	0.1	0.3	1.8	0.8
Oxygen uptake in coal	-	-	-	-	-	3.9
Utilities						
Electricity, % of input	0.04	0.08	0.04	0.6	0.5	1.7
Steam, % of input	0	0	0	4.5	6.2	9.2
Natural gas, % of input	0	0	0	0.02	0	0
Diesel fuel, % of input	0.02	0.02	0.02	0	0	0
Total	9.6	9.3	9.4	5.4	8.5	16.8

Basis

2,000-MW utility power plant, 5,500 hr/yr operation. 9,500 Btu/kWh design heat rate, 5% sulfur coal.

The Kennecott process also has a small coal Btu usage because a portion of the coal chemical structure is altered during the cleaning process. This energy aids in holding the reaction temperature at the desired level, thereby replacing an equivalent amount of energy in the form of steam. In addition, the coal product from the Kennecott process has an oxygen uptake resulting in an additional Btu loss.

The KVB process utilizes a reaction at atmospheric pressure and at relatively low temperatures. As a result, the 5.4% total energy usage for the KVB process is significantly better than for the other processes.

GROUND-TO-GROUND FGD STUDY

In previous years EPA, TVA, and others have investigated the design and economics of limestone, lime, and magnesia scrubbing processes. Since these earlier studies, technical and operating information on these systems has greatly increased. Many full-scale applications of the limestone and lime systems are now in operation and the magnesia process has also been evaluated to a lesser extent in full-scale operation. This study is a continuation of the earlier design and economic evaluations but incorporates the most recently available commercial technology.

The design of the MgO process has been updated to include recent commercial technology. Provision for chloride removal has been added in the scrubbing system. In the regeneration area process refinements include substantial use of pneumatic conveyors instead of belts, the use of a rotary dryer instead of a fluidized-bed dryer, redesigned slaking and calcining systems, a simplified slurry processing system, and additional heat-recovery equipment.

In addition, a special energy requirement assessment of the three systems is included. Energy costs are expected to increase more rapidly than other costs associated with FGD processes. Increases in energy costs relative to other FGD costs could radically change the comparative economics of energy-intensive processes and processes with low energy requirements. The energy requirement assessment is a ground-to-ground study including all requirements for raw material preparation and waste disposal as well as process energy usage.

PROCESS DESCRIPTIONS

The FGD systems are assumed to be installed downstream from the power plant particulate-control units, beginning with a single plenum which collects all the power plant flue gas. This plenum supplies four parallel trains of FGD equipment which vent to the stack plenum. Particulate control to meet NSPS is not included in the FGD costs. Each train is equipped with a forced-draft booster fan to compensate for pressure loss in the FGD system and with indirect-steam flue gas reheating to 175°F. A presaturator is included upstream from the scrubber to reduce the flue gas temperature from 300°F to 127°F and a mist eliminator is included downstream from the scrubber to reduce the flue gas moisture to 0.1%. A single raw materials and feed preparation area and a single waste disposal system serve all four FGD trains.

Limestone Process

The widely used limestone process, Figure 11, scrubs the flue gas with a slurry of finely ground limestone, forming hydrated calcium-sulfur salts which are discarded as a waste sludge. Purchased limestone is crushed and wet ground to form the absorbent feed slurry. The relatively low reactivity of limestone requires a stoichiometric ratio of 1.3 moles of $CaCO_3$ to 1.0 mole of sulfur removed.

A countercurrent mobile-bed scrubber is used. A 15% solids slurry is circulated through the scrubber countercurrent to the flue gas and through an external loop where makeup absorbent slurry is added and a purge stream containing $CaSO_3 \cdot 1/2H_2O$, $CaSO_4 \cdot 2H_2O$, and unreacted limestone is removed. The purge stream is pumped to an earthen-diked, clay-lined disposal pond where it settles to a sludge of about 40% solids. The supernate is returned to the scrubber system.

Lime Process

The lime process, Figure 12, is similar to the limestone process except for details of absorbent preparation and process chemistry. Pebble lime is slaked, forming a slurry of $Ca(OH)_2$ which is used as absorbent feed. Two cases are evaluated, one in which purchased lime is used and a second in which an onsite calciner, Figure 13, is included to produce lime from limestone. The higher reactivity of lime as compared to limestone permits a stoichiometric ratio of 1.05 moles of $Ca(OH)_2$ to 1.00 mole of SO_x removed. This results in a small reduction in slurry-handling equipment size and pond capacity compared to the limestone process.

A countercurrent mobile-bed absorber is used. The 15% solids slurry is circulated, absorbent feed added, and a purge stream removed in the same manner as in the limestone process. The purge stream, consisting of $CaSO_3 \cdot 1/2H_2O$ and $CaSO_4 \cdot 2H_2O$ with a small quantity of unreacted $Ca(OH)_2$, is pumped to an earthen-diked, clay-lined pond where it settles to a sludge of about 40% solids. Supernate is returned to the scrubber system.

Magnesia Process

The magnesia process, Figure 14, consists of a wet-scrubbing system using MgO as the absorbent. The use of MgO provides a rapid absorption reaction and a low scaling potential, allowing a wide latitude of design and operating conditions. The cost of MgO, however, necessitates a regeneration process. In this study the absorber reactants are calcined to produce MgO and SO_2 and the SO_2 is processed to H_2SO_4.

A countercurrent mobile-bed scrubber is used. The feed is an $Mg(OH)_2$ slurry prepared by slaking MgO in water. The presaturator bottoms are discarded in the ash pit to control residual fly ash and

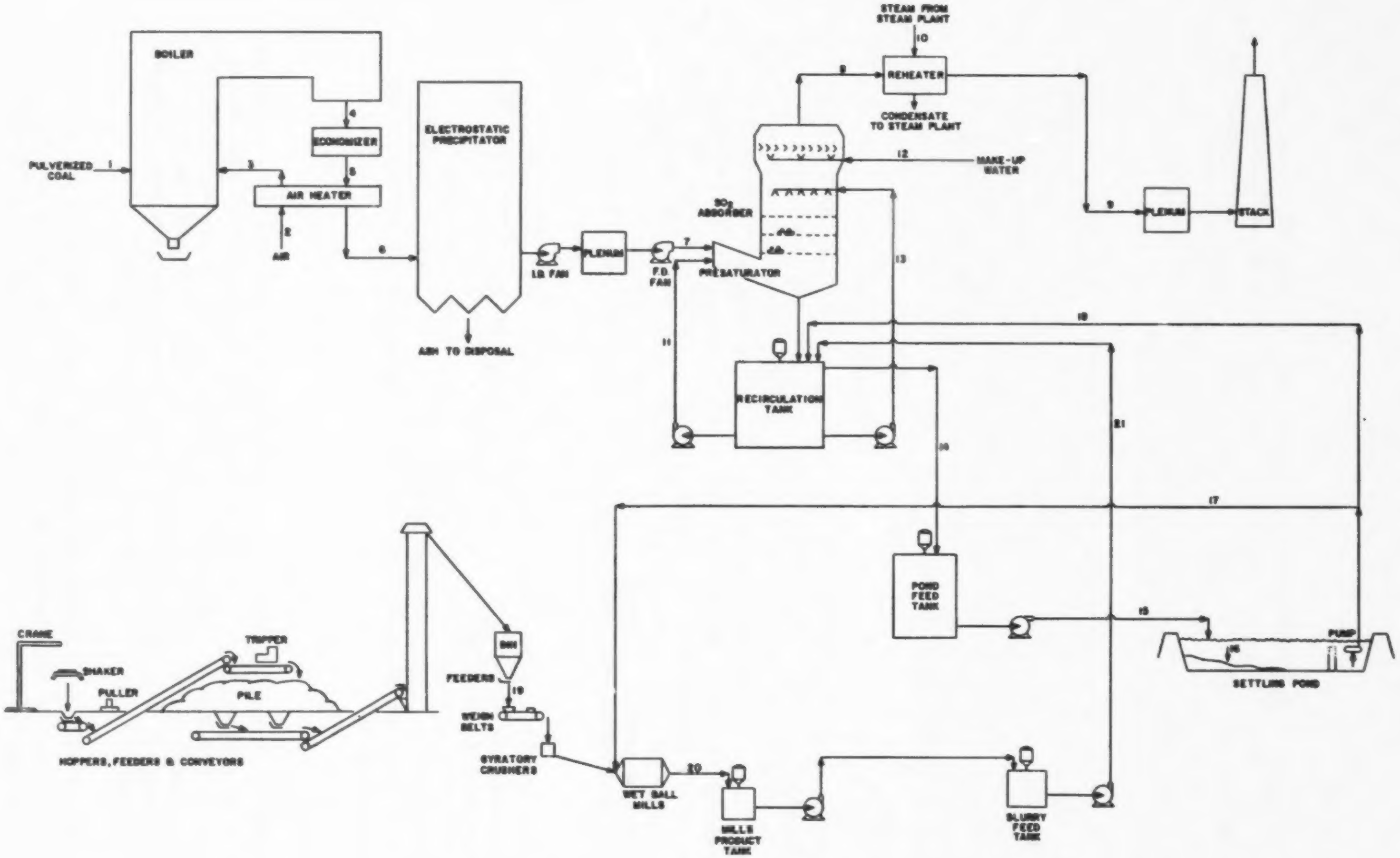
BOILER
PULVERIZED COAL
1
2
AIR
3
4
ECONOMIZER
5
AIR HEATER
6
ELECTROSTATIC PRECIPITATOR
ASH TO DISPOSAL
I.D. FAN
PLENUM
F.D. FAN
7
SO_2 ABSORBER
PRESATURATOR
8
STEAM FROM STEAM PLANT
10
REHEATER
CONDENSATE TO STEAM PLANT
12
MAKE-UP WATER
9
PLENUM
STACK
11
13
RECIRCULATION TANK
14
18
21
17
POND FEED TANK
15
16
PUMP
SETTLING POND
CRANE
SHAKER
PULLER
TRIPPER
PILE
HOPPERS, FEEDERS & CONVEYORS
BIN
FEEDERS
19
WEIGH BELTS
GYRATORY CRUSHERS
WET BALL MILLS
20
MILLS PRODUCT TANK
SLURRY FEED TANK

Figure 11. Limestone slurry process.

Flow diagram.

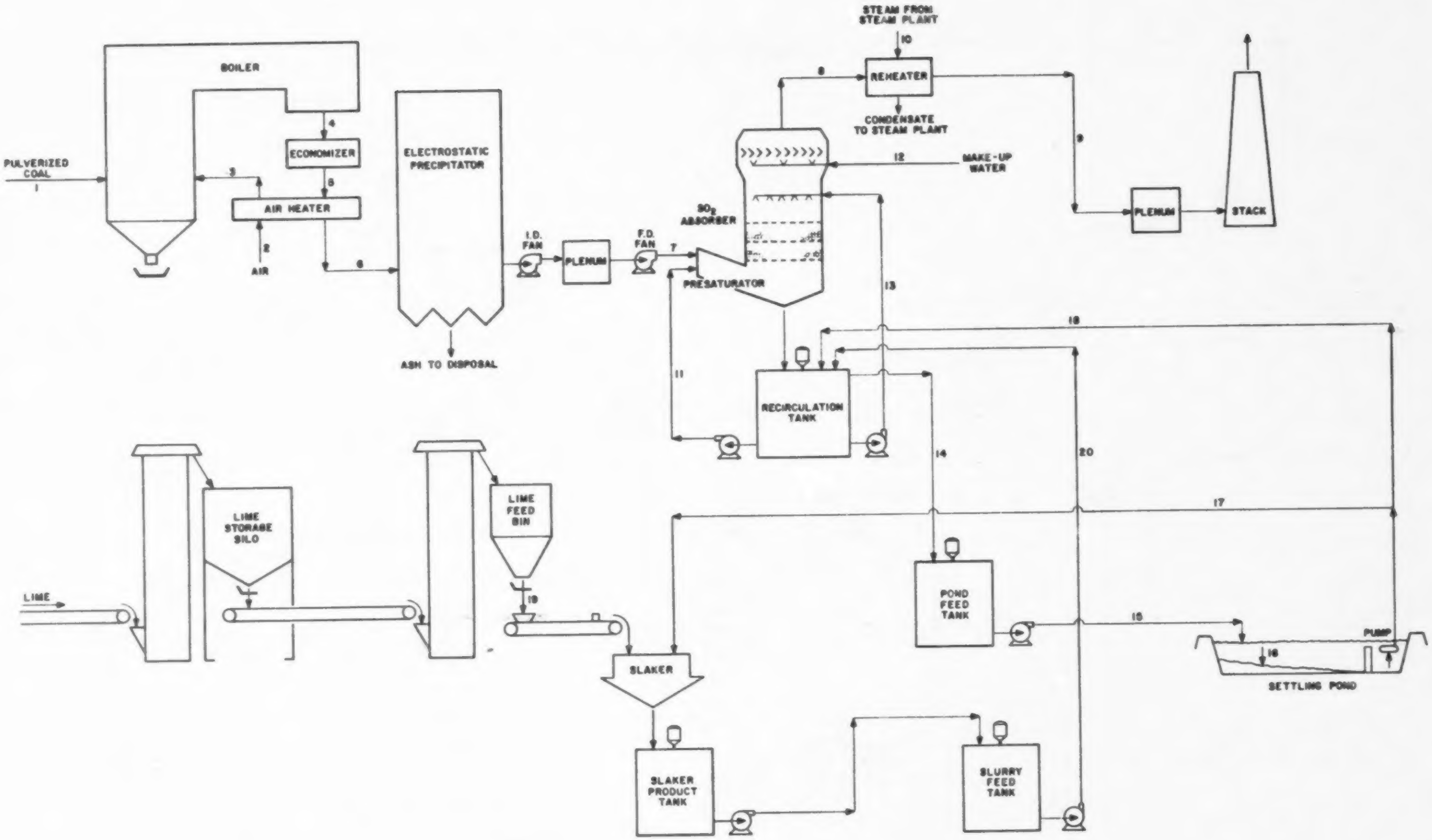
STEAM FROM STEAM PLANT
10
REHEATER
8
CONDENSATE TO STEAM PLANT
9
BOILER
4
ECONOMIZER
PULVERIZED COAL
1
3
5
AIR HEATER
2
AIR
6
ELECTROSTATIC PRECIPITATOR
ASH TO DISPOSAL
I.D. FAN
PLENUM
F.D. FAN
7
SO_2 ABSORBER
PRESATURATOR
12
MAKE-UP WATER
13
PLENUM
STACK
18
11
RECIRCULATION TANK
14
20
17
LIME
LIME STORAGE SILO
LIME FEED BIN
19
SLAKER
POND FEED TANK
15
PUMP
16
SETTLING POND
SLAKER PRODUCT TANK
SLURRY FEED TANK

Figure 12. Lime slurry process.

Flow diagram.

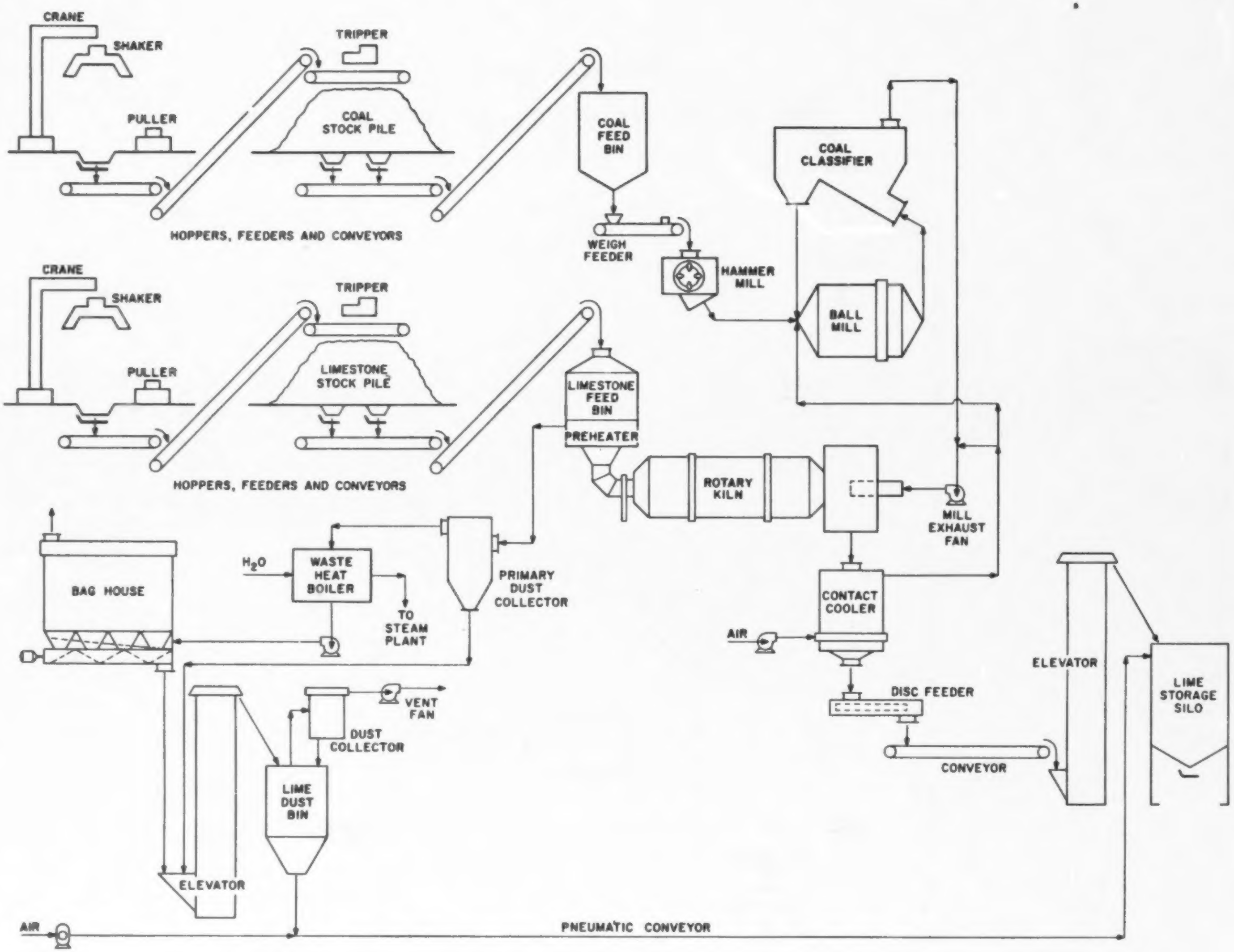
CRANE
SHAKER
PULLER
TRIPPER
COAL STOCK PILE
HOPPERS, FEEDERS AND CONVEYORS
COAL FEED BIN
WEIGH FEEDER
HAMMER MILL
COAL CLASSIFIER
BALL MILL
CRANE
SHAKER
PULLER
TRIPPER
LIMESTONE STOCK PILE
HOPPERS, FEEDERS AND CONVEYORS
LIMESTONE FEED BIN
PREHEATER
ROTARY KILN
MILL EXHAUST FAN
BAG HOUSE
H_2O
WASTE HEAT BOILER
TO STEAM PLANT
PRIMARY DUST COLLECTOR
CONTACT COOLER
AIR
ELEVATOR
LIME STORAGE SILO
DISC FEEDER
CONVEYOR
VENT FAN
DUST COLLECTOR
LIME DUST BIN
ELEVATOR
AIR
PNEUMATIC CONVEYOR

Figure 13. Lime calcination process.

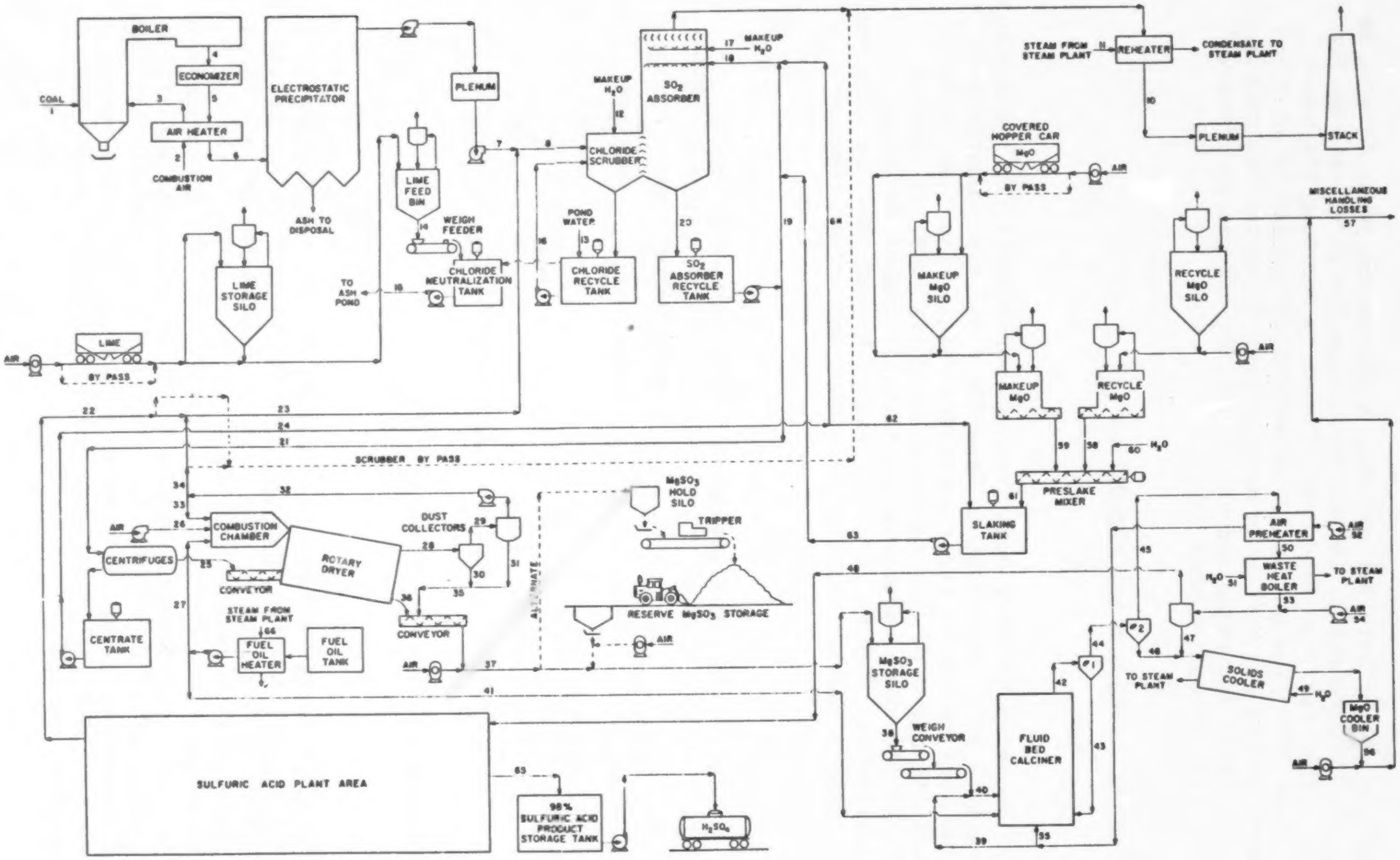
BOILER
COAL
ECONOMIZER
AIR HEATER
COMBUSTION AIR
ELECTROSTATIC PRECIPITATOR
ASH TO DISPOSAL
PLENUM
LIME FEED BIN
WEIGH FEEDER
CHLORIDE NEUTRALIZATION TANK
TO ASH POND
LIME STORAGE SILO
LIME
BY PASS
AIR
MAKEUP H_2O
CHLORIDE SCRUBBER
SO_2 ABSORBER
POND WATER
CHLORIDE RECYCLE TANK
SO_2 ABSORBER RECYCLE TANK
STEAM FROM STEAM PLANT
REHEATER
CONDENSATE TO STEAM PLANT
PLENUM
STACK
COVERED HOPPER CAR
MgO
MAKEUP MgO SILO
RECYCLE MgO SILO
MISCELLANEOUS HANDLING LOSSES
MAKEUP MgO
RECYCLE MgO
PRESLAKE MIXER
H_2O
SLAKING TANK
SCRUBBER BY PASS
COMBUSTION CHAMBER
CENTRIFUGES
ROTARY DRYER
CONVEYOR
DUST COLLECTORS
CENTRATE TANK
STEAM FROM STEAM PLANT
FUEL OIL HEATER
FUEL OIL TANK
ALTERNATE
$MgSO_3$ HOLD SILO
TRIPPER
RESERVE $MgSO_3$ STORAGE
AIR PREHEATER
WASTE HEAT BOILER
TO STEAM PLANT
$MgSO_3$ STORAGE SILO
WEIGH CONVEYOR
FLUID BED CALCINER
SOLIDS COOLER
TO STEAM PLANT
MgO COOLER BIN
SULFURIC ACID PLANT AREA
98% SULFURIC ACID PRODUCT STORAGE TANK
H_2SO_4

Figure 14. Magnesia slurry-regeneration process.

Flow diagram.

chloride buildup in the scrubber system. A 15% solids purge stream, in which $MgSO_3 \cdot 6H_2O$ predominates, is withdrawn from the scrubber loop for regeneration. The slurry is first centrifuged to decrease the water content to about 15%. The solids are dried and dehydrated in an oil-fired rotary kiln and decomposed in an oil-fired fluidized-bed calciner. The MgO is collected for reuse and the 10-15% SO_2 off-gas is converted to H_2SO_4 in a single-contact acid plant.

RESULTS OF GROUND-TO-GROUND FGD EVALUATION

Economics

Capital investments and annual revenue requirements for the processes evaluated in this study are shown in Appendix B and are summarized in Table 4. The ranking of the three processes remains the same as in previous evaluations, with lime scrubbing slightly lower in capital investment than limestone scrubbing and slightly higher in annual revenue requirements. The MgO process remains higher than the lime and limestone processes in both capital investment and annual revenue requirements.

TABLE 4. FGD CAPITAL INVESTMENT AND ANNUAL REVENUE REQUIREMENTS

Process	Capital investment[b] $	Capital investment[b] $/kW	Annual revenue requirements[a] $/yr	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton sulfur removed
Limestone	48,948,000	97.90	14,375,300	4.11	9.58	0.46	411
Lime - calcination	53,859,000	107.72	15,531,200	4.44	10.35	0.49	444
Lime	45,319,000	90.64	14,890,500	4.25	9.93	0.47	425
Magnesia	70,293,000	140.59	18,325,000[c]	5.24	12.22	0.58	524

a. 1980 dollars.
b. 1979 dollars.
c. Includes $3,412,100 sulfuric acid sales credit.

In comparison to the results reported in 1976 (G. G. McGlamery, et al., *Flue Gas Desulfurization Economics*) at the Sixth Flue Gas Desulfurization Symposium in New Orleans, the capital investments for all three processes are greatly increased in this study, particularly for the MgO process. This comparison can be seen in Figure 15. The differences in these results are from premise changes, design factors, and inflation. The MgO process capital investment has increased more dramatically than the lime or limestone, primarily because of changes in technology. Annual revenue requirements show much smaller increases between the 1976 and the 1979 results.

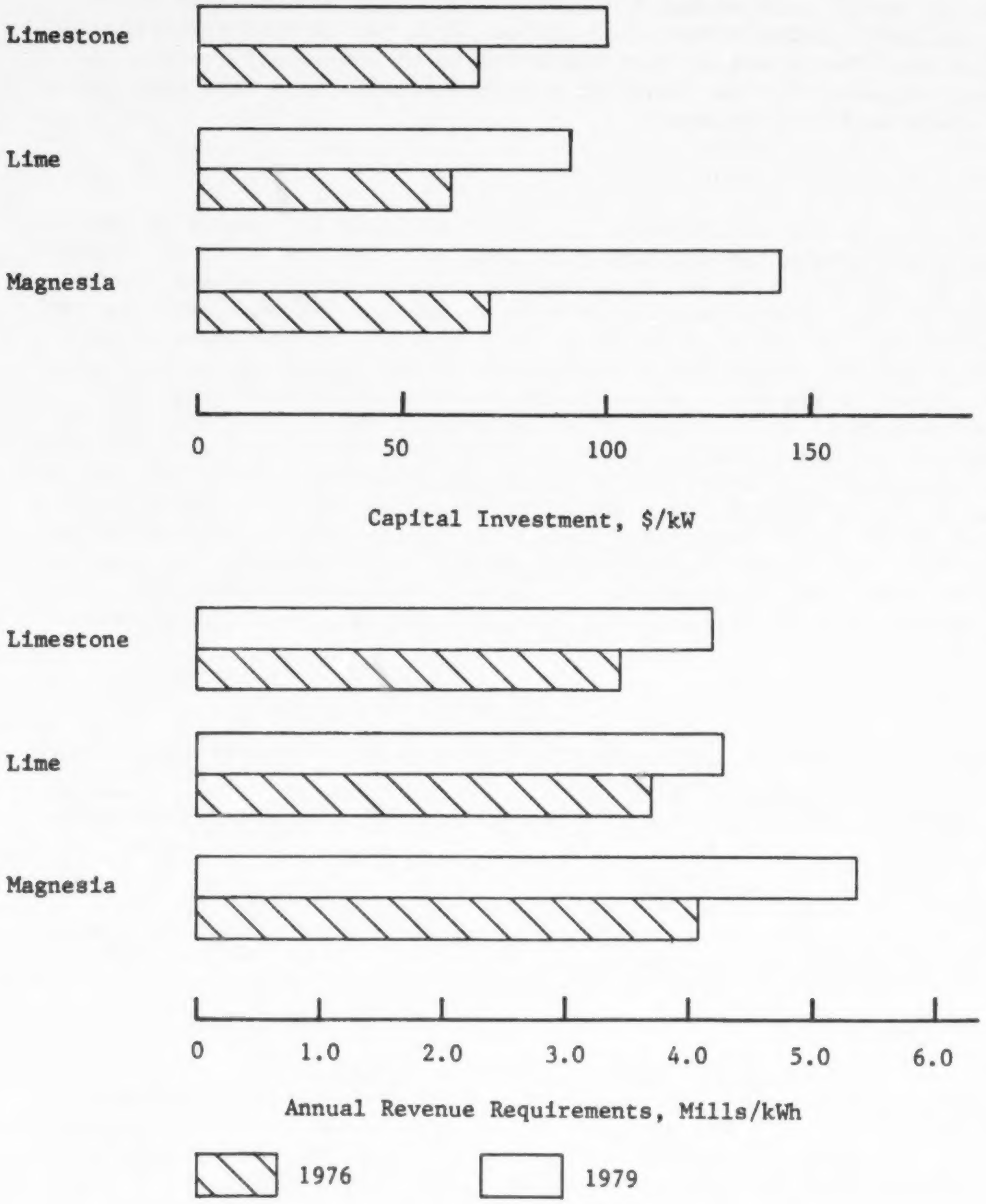

Figure 15. Comparison of 1976 and 1979 evaluations.

One interesting result appears when comparing lime scrubbing and limestone onsite calcination facilities with the purchased lime case. Data available indicate that lime can be purchased in most situations at a price less than a utility can manufacture its own supply. This, of course, suggests that the costs of a larger supplier are less than those of a small onsite operation.

Energy

Process energy requirements are shown in Table 5. Ground-to-ground energy requirements for the same processes are shown in Table 6. Figures 16 and 17 show the same data graphically. The ground-to-ground energy results include energy required to mine or produce and transport the raw materials required as well as the process energy. On the basis chosen raw material production energy has little effect on the overall energy requirements of the processes. Limestone scrubbing ground-to-ground energy is only slightly higher than the process energy, illustrating the relatively low energy requirements of limestone quarrying and transportation. The use of lime instead of limestone scrubbing reduces FGD process energy slightly but increases ground-to-ground energy requirements because of the large heat requirements for calcination. Interestingly, onsite lime calcination consumes more energy than the energy represented by commercial lime. Electricity, transportaion, and calciner fuel requirements are all higher for the relatively small onsite calciner.

TABLE 5. FGD PROCESS ENERGY REQUIREMENTS

Process	Electricity		Steam		Oil, MBtu/hr	Coal, MBtu/hr	Heat credit, MBtu/hr	Total equivalent energy consumption,[a] % of input energy
	MBtu/hr	kW	Reheat, MBtu/hr	Process, MBtu/hr				
Limestone	68.9	7,655	70.0	-	-	-	-	3.3
Lime - calcination	65.9	7,326	70.0	-	-	58.9	(3.6)	4.4
Lime	60.4	6,715	70.0	-	-	-	-	3.1
Magnesia	67.2	7,468	71.3	0.6	117.7	-	(11.9)	5.6

a. Based on a 500-MW boiler efficiency of 90% for generation of steam and a gross heat rate of 9,000 Btu/kWh for generation of electricity.

TABLE 6. GROUND-TO-GROUND FGD ENERGY REQUIREMENTS

Process	Electricity		Steam		Oil, MBtu/hr	Natural gas, MBtu/hr	Coal, MBtu/hr	Heat credit, MBtu/hr	Total equivalent energy consumption,[a] % of input energy
	MBtu/hr	kW	Reheat, MBtu/hr	Process, MBtu/hr					
Limestone	69.1	7,700	70.0	-	8.9[b]	-	-	-	3.5
Lime - calcination	66.2	7,400	70.0	-	7.3[b]	-	58.9	(3.6)	4.6
Lime	62.8	7,700	70.0	-	4.3[b]	-	50.0	-	4.3
Magnesia - magnesite	72.0	8,900	71.3	0.6	119.0[c]	-	-	(11.9)	5.7
Magnesia - seawater	75.2	9,800	71.3	0.6	121.3[c]	1.7	-	(11.9)	5.9

a. Based on a 500-MW boiler efficiency of 90% for generation of steam and a gross heat rate of 9,000 Btu/kWh for generation of electricity. Based on 1.2 lb SO_2/MBtu heat input allowable emission.

b. All limestone quarry and offsite lime processing plants are assumed to be 100 miles from the Chicago area FGD plant.

c. Oil energy includes transporting MgO from Gabbs, Nevada, for magnesite and Port St. Joe, Florida, for seawater to a Chicago area FGD plant.

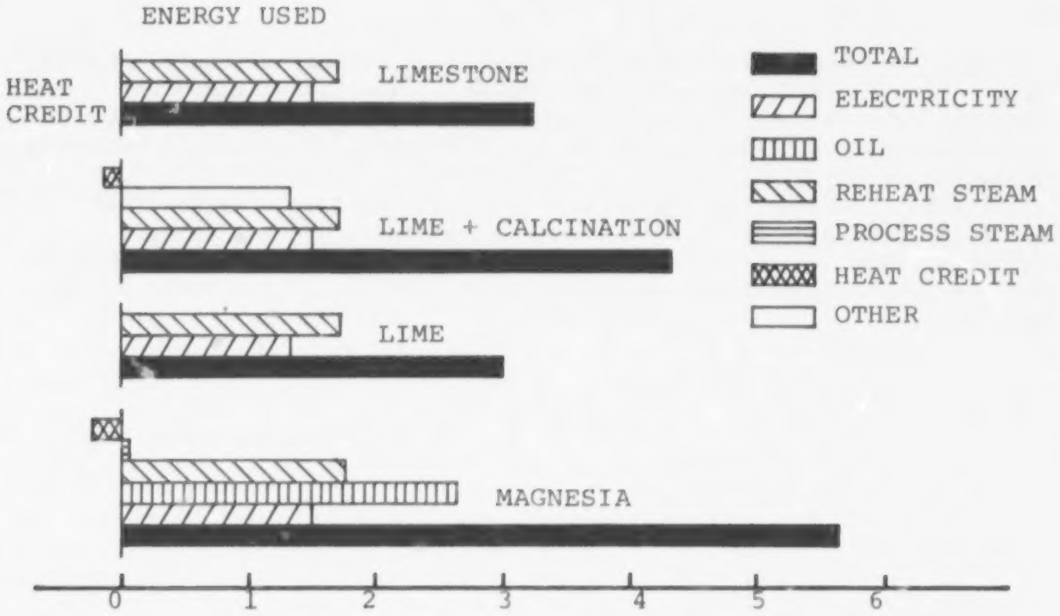

FIGURE 16. PROCESS ENERGY REQUIREMENTS.

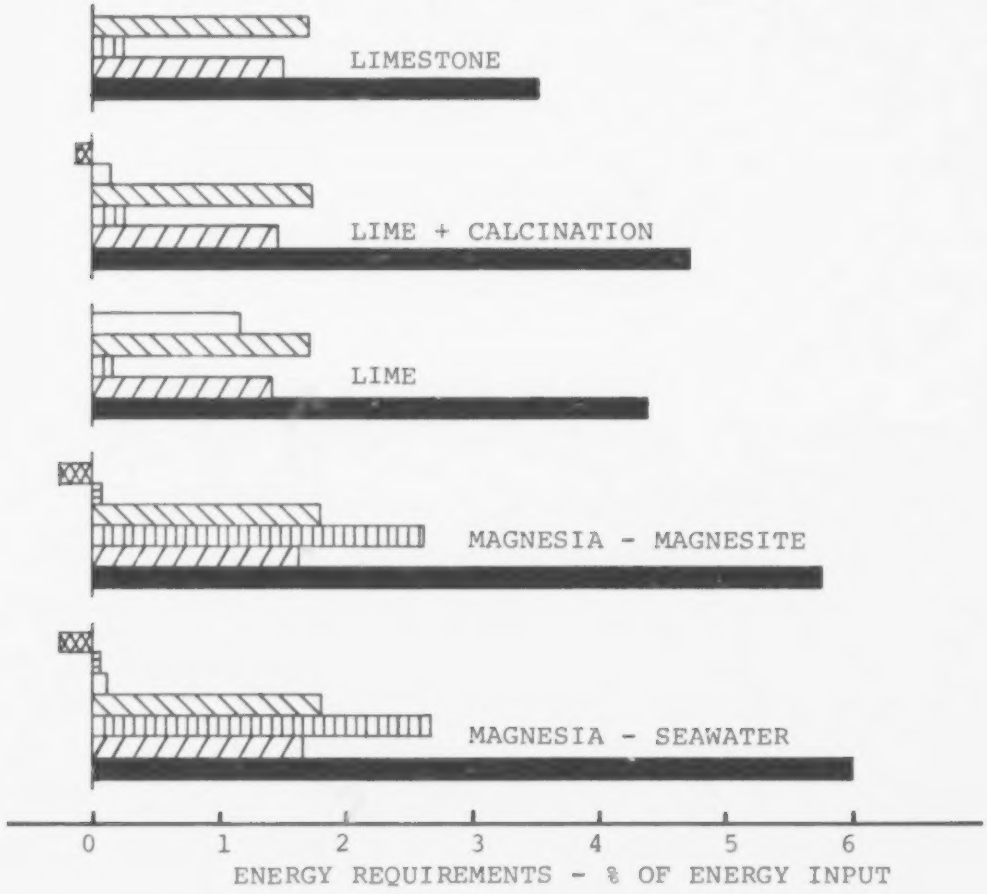

Figure 17. Ground-to-ground energy requirements.

The ground-to-ground magnesia energy requirements are only slightly higher than the process energy requirements. The use of magnesia produced by treatment of dolomite with seawater results in a slightly higher energy requirement than the use of magnesia from natural magnesite deposits. Both types of magnesia require large amounts of energy to produce but, because the magnesia is regenerated, a relatively small amount is used and the overall energy requirements are not greatly affected. Lime, in contrast, requires considerably less energy to produce but the larger quantities used result in a substantially greater energy use.

REEVALUATION OF THE WELLMAN-LORD FGD PROCESS

The Wellman-Lord FGD system developed by Davy Powergas, Inc., has been used in several industrial and utility applications and has undergone some changes since its previous economic evaluation by TVA in 1974. It is now in operation at U.S. coal-fired utility power plants. Also, as a replacement for natural gas, new coal-reduction options are being developed for processing the concentrated SO_2 from the Wellman-Lord unit. This evaluation incorporates the latest Wellman-Lord scrubbing technology and two SO_2 conversion processes, the Allied Chemical coal-reduction process which produces sulfur, and a conventional H_2SO_4 plant.

Design changes in the Wellman-Lord scrubbing process represent both continuing development and experience on U.S. coal-fired boilers. An improved raw material wet storage system is now used. Chloride and SO_3 removal and neutralization are provided prior to SO_2 removal. The regeneration system is substantially revised to correspond to current design practice. Filters have been added to control ash buildup, tank capacity has been increased, and a high-temperature Na_2SO_4 crystallization system based on a revised oxidation rate is utilized. No longer is an antioxidant used. Double-effect evaporators replace the previous single-effect evaporators as a result of current energy conservation practice.

PROCESS DESCRIPTIONS

Wellman-Lord Process

The Wellman-Lord process, shown in Figure 18, is a regeneration wet-scrubbing process using a solution of Na_2SO_3 as the absorbent. Regeneration of the absorbent produces a concentrated SO_2 stream which can be processed to either sulfur or H_2SO_4.

A spray-type presaturator using recycled water is used for chloride, SO_3, and residual fly ash control. Presaturator bottoms are discarded in the power plant ash pond. The absorber consists of a countercurrent-flow, three-stage valve tray unit with separate recirculation in each stage. The scrubber effluent containing the reaction products, which are primarily $NaHSO_3$ and Na_2SO_4, is processed to remove Na_2SO_4 and regenerate Na_2SO_3 and SO_2. Sodium losses are made up by addition of Na_2CO_3 to the regenerated absorbent.

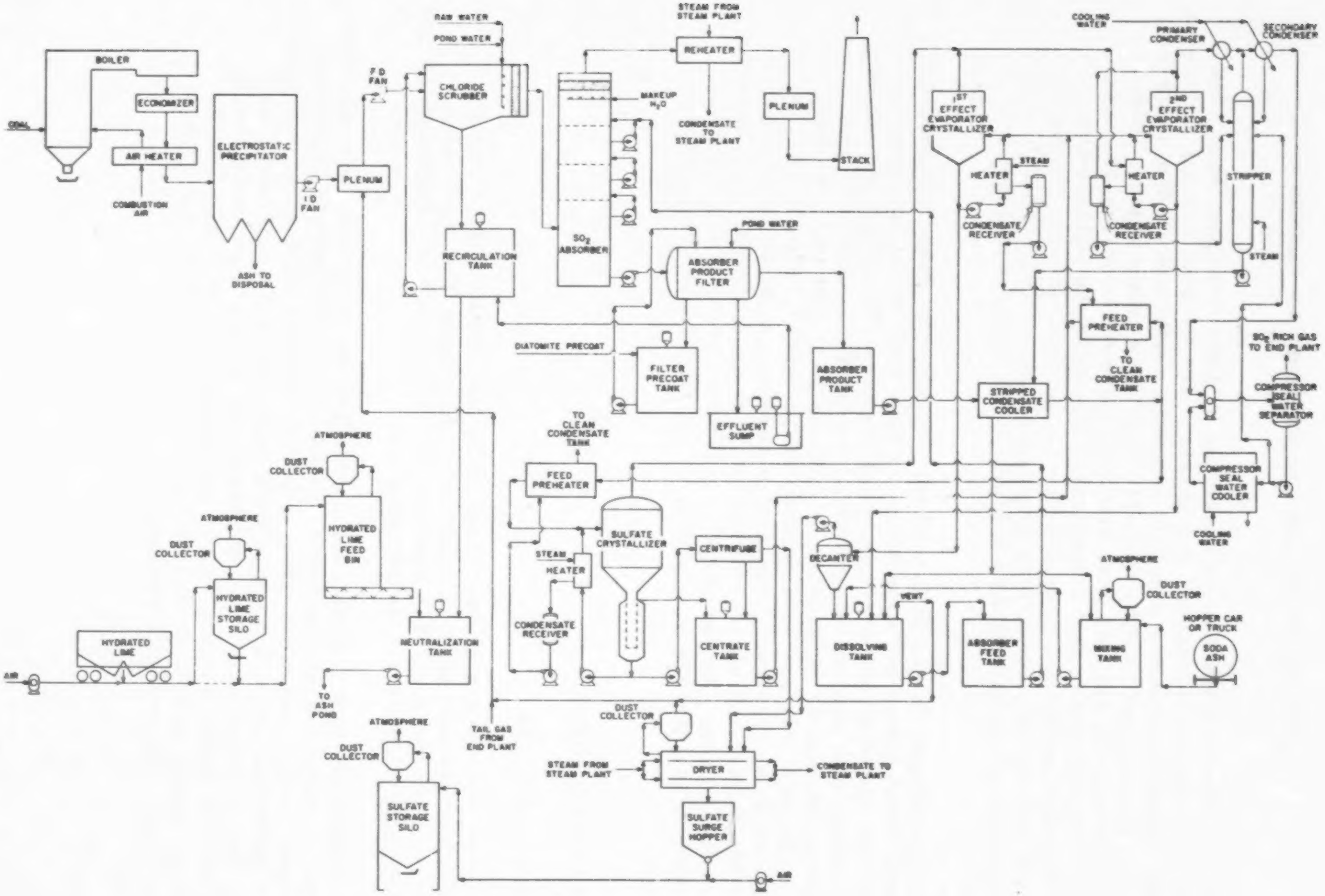
COAL
BOILER
ECONOMIZER
AIR HEATER
COMBUSTION AIR
ELECTROSTATIC PRECIPITATOR
ASH TO DISPOSAL
I D FAN
PLENUM
F D FAN
RAW WATER
POND WATER
CHLORIDE SCRUBBER
RECIRCULATION TANK
MAKEUP H_2O
SO_2 ABSORBER
STEAM FROM STEAM PLANT
REHEATER
CONDENSATE TO STEAM PLANT
PLENUM
STACK
POND WATER
ABSORBER PRODUCT FILTER
DIATOMITE PRECOAT
FILTER PRECOAT TANK
EFFLUENT SUMP
ABSORBER PRODUCT TANK
COOLING WATER
PRIMARY CONDENSER
SECONDARY CONDENSER
1ST EFFECT EVAPORATOR CRYSTALLIZER
2ND EFFECT EVAPORATOR CRYSTALLIZER
HEATER
STEAM
CONDENSATE RECEIVER
HEATER
CONDENSATE RECEIVER
STRIPPER
STEAM
FEED PREHEATER
TO CLEAN CONDENSATE TANK
STRIPPED CONDENSATE COOLER
SO_2 RICH GAS TO END PLANT
COMPRESSOR SEAL WATER SEPARATOR
COMPRESSOR SEAL WATER COOLER
COOLING WATER
ATMOSPHERE
DUST COLLECTOR
HYDRATED LIME FEED BIN
ATMOSPHERE
DUST COLLECTOR
HYDRATED LIME STORAGE SILO
HYDRATED LIME
AIR
NEUTRALIZATION TANK
TO ASH POND
TO CLEAN CONDENSATE TANK
FEED PREHEATER
STEAM
HEATER
CONDENSATE RECEIVER
SULFATE CRYSTALLIZER
CENTRIFUGE
CENTRATE TANK
DECANTER
VENT
DISSOLVING TANK
ABSORBER FEED TANK
ATMOSPHERE
DUST COLLECTOR
MIXING TANK
HOPPER CAR OR TRUCK
SODA ASH
TAIL GAS FROM END PLANT
ATMOSPHERE
DUST COLLECTOR
SULFATE STORAGE SILO
DUST COLLECTOR
STEAM FROM STEAM PLANT
DRYER
CONDENSATE TO STEAM PLANT
SULFATE SURGE HOPPER
AIR

Figure 18. Wellman-Lord SO_2 recovery process.

Flow diagram.

A portion of the scrubber effluent is processed to remove Na_2SO_4 by evaporation and selective crystallization in a steam-heated, forced-circulation evaporator serving all four scrubber trains. The clear overflow, enriched in $NaHSO_3$, is returned to the regeneration area. The bottoms, consisting of a slurry enriched in Na_2SO_4 crystals, are centrifuged to produce a solid containing about two-thirds Na_2SO_4 and one-third Na_2SO_3. The centrate is returned to the regeneration area; the solids are dried in a steam-heated dryer and conveyed to a storage silo for sale or discard. There is a potential market available in the paper industry for this material.

The regeneration system consists of two trains of double-effect, forced-circulation evaporators. Scrubber effluent, combined with liquid from the sulfate removal process, is heated and 60% is pumped to the first-effect evaporators and 40% is pumped to the second-effect evaporators. The first effect is steam heated; the second effect is heated by combined first-effect vapor and sulfate crystallizer vapor. Some $Na_2S_2O_3$ formed in the first-effect evaporator is removed by a purge stream. Evaporator bottoms, consisting primarily of Na_2SO_3 are returned to the absorbent system. Evaporator and stripper overhead vapor, containing H_2O and SO_2, is dried and the SO_2 is sent to a processing plant. SO_2-bearing condensate from the second-effect evaporator heater, the condensers, and the compressor is steam stripped and returned to the absorber system.

The SO_2 can be processed to either sulfur or H_2SO_4 by several methods. In this study an Allied Chemical Corporation process for producing sulfur and a single-contact, single-absorption acid plant to produce H_2SO_4 are evaluated.

Allied Chemical Process

A proprietary process, shown in Figure 19, developed by Allied Chemical Corporation reduces the SO_2 to sulfur using coal. Powdered coal is injected into a reactor containing a bed of inert material and the bed is fluidized with heated SO_2 and air. Air is added because some additional oxidation of coal is needed to maintain reaction temperature. About 70% of the SO_2 is reduced to sulfur which is condensed from the off-gas. The remaining off-gas containing SO_2 and H_2S is passed through a Claus-type catalytic converter to recover additional sulfur. The Claus converter off-gas is oxidized and recycled to the SO_2 absorber.

Sulfuric Acid Plant

For the H_2SO_4 alternative the SO_2 is converted to H_2SO_4 in a single-contact, single-absorption acid plant. A single-contact plant is used for economy and the tail gas containing unreacted SO_2 is returned to the scrubber. A flowsheet of the acid plant is shown in Figure 20.

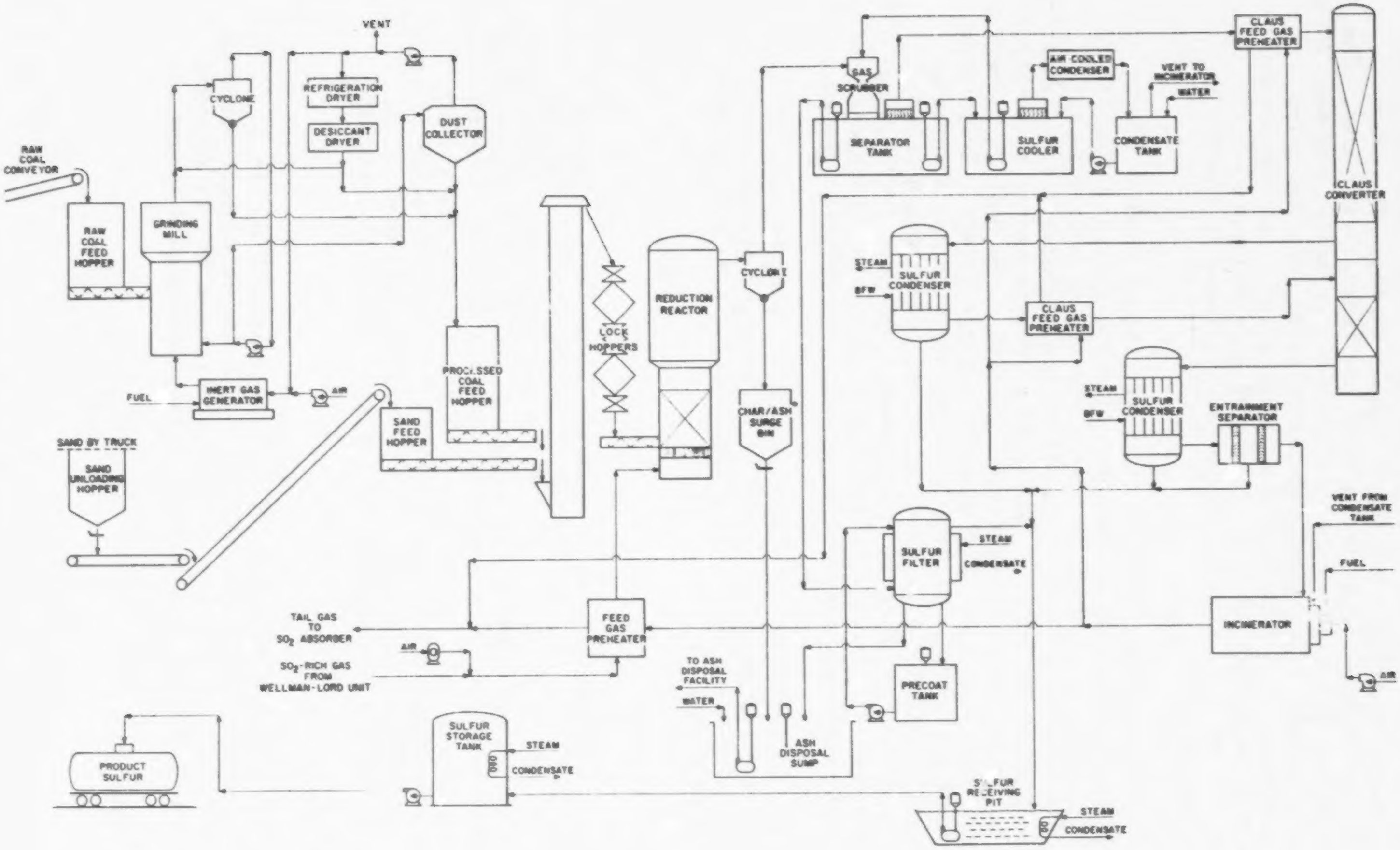
VENT
RAW COAL CONVEYOR
RAW COAL FEED HOPPER
GRINDING MILL
CYCLONE
REFRIGERATION DRYER
DESICCANT DRYER
DUST COLLECTOR
INERT GAS GENERATOR
FUEL
AIR
SAND BY TRUCK
SAND UNLOADING HOPPER
SAND FEED HOPPER
PROCESSED COAL FEED HOPPER
LOCK HOPPERS
REDUCTION REACTOR
CYCLONE
CHAR/ASH SURGE BIN
GAS SCRUBBER
SEPARATOR TANK
SULFUR COOLER
AIR COOLED CONDENSER
CONDENSATE TANK
VENT TO INCINERATOR
WATER
CLAUS FEED GAS PREHEATER
CLAUS CONVERTER
STEAM
BFW
SULFUR CONDENSER
CLAUS FEED GAS PREHEATER
STEAM
BFW
SULFUR CONDENSER
ENTRAINMENT SEPARATOR
VENT FROM CONDENSATE TANK
FUEL
INCINERATOR
AIR
SULFUR FILTER
STEAM
CONDENSATE
PRECOAT TANK
TAIL GAS TO SO2 ABSORBER
AIR
SO2-RICH GAS FROM WELLMAN-LORD UNIT
FEED GAS PREHEATER
TO ASH DISPOSAL FACILITY
WATER
ASH DISPOSAL SUMP
PRODUCT SULFUR
SULFUR STORAGE TANK
STEAM
CONDENSATE
SULFUR RECEIVING PIT
STEAM
CONDENSATE

Figure 19. Allied Chemical coal/SO_2 reduction technology.

Flow diagram.

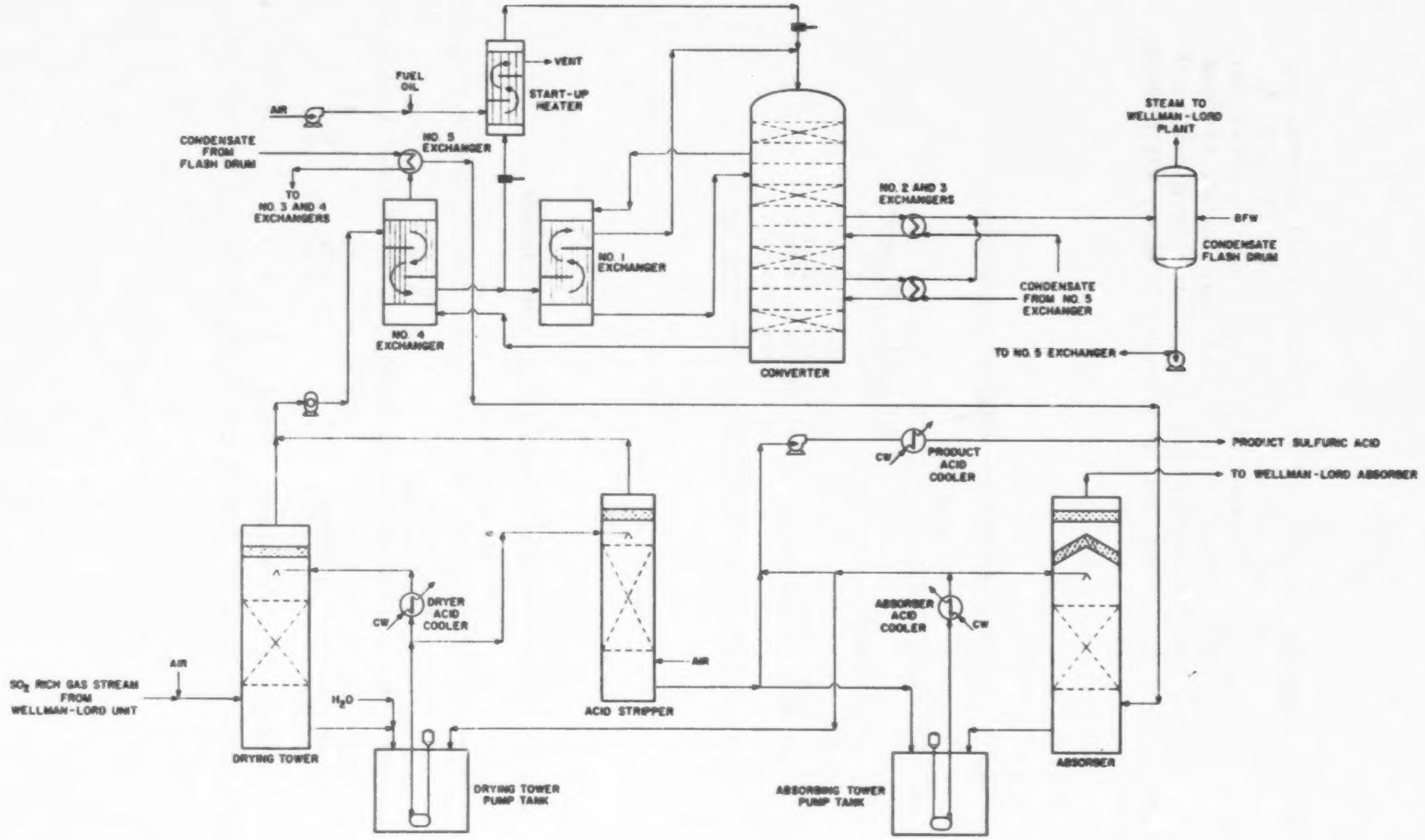
AIR
FUEL OIL
VENT
START-UP HEATER
CONDENSATE FROM FLASH DRUM
NO. 5 EXCHANGER
TO NO. 3 AND 4 EXCHANGERS
NO. 4 EXCHANGER
NO. 1 EXCHANGER
CONVERTER
NO. 2 AND 3 EXCHANGERS
CONDENSATE FROM NO. 5 EXCHANGER
STEAM TO WELLMAN-LORD PLANT
BFW
CONDENSATE FLASH DRUM
TO NO. 5 EXCHANGER
PRODUCT SULFURIC ACID
TO WELLMAN-LORD ABSORBER
CW
PRODUCT ACID COOLER
SO_2 RICH GAS STREAM FROM WELLMAN-LORD UNIT
AIR
DRYING TOWER
CW
DRYER ACID COOLER
H_2O
DRYING TOWER PUMP TANK
AIR
ACID STRIPPER
ABSORBER ACID COOLER
CW
ABSORBING TOWER PUMP TANK
ABSORBER

Figure 20. Sulfuric acid plant.

Flow diagram.

RESULTS

The bottom line economic results of the Wellman-Lord process are given in Tables 7 and 8 along with those for the citrate (from R. L. Torstrick et al.) and magnesia processes which have been repeated for easy comparison. The summary tables of investment and annual revenue requirements for the Wellman-Lord scrubbing system with both acid and sulfur production options are presented in Appendix B along with those for other processes.

TABLE 7. SUMMARY OF RECOVERY FGD INVESTMENT REQUIREMENTS

1979 DOLLARS

Process	Total capital investment $	 $/kW
Magnesia	70,293,000	140.59
Wellman Lord - H_2SO_4	71,448,000	142.90
Wellman Lord - Allied Chemical	74,190,000	148.38
Citrate	74,918,000	149.84

TABLE 8. SUMMARY OF RECOVERY FGD ANNUAL REVENUE REQUIREMENTS

1980 DOLLARS

Process	Gross average annual revenue requirements	Net average annual revenue requirements	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton sulfur removed
Magnesia	21,025,100	18,325,000	5.24	12.22	0.58	524
Wellman Lord - H_2SO_4	21,752,800	19,058,800	5.44	12.71	0.61	545
Wellman Lord - Allied Chemical	23,151,900	21,478,400	6.14	14.32	0.68	614
Citrate	24,820,800	23,298,000	6.58	15.35	0.73	666

Producing sulfuric acid by the Wellman-Lord system requires about 4% less capital investment and 15% less annual revenue than the Allied sulfur production option. When the Wellman-Lord acid option is compared to the magnesia process, it requires about 2% more capital investment and 4% more annual revenue. Comparing sulfur options, the Wellman-Lord - Allied system requires about 1% less capital investment and 6% less annual revenue than the citrate process. It should be stated, however, that the costs derived for this paper are preliminary and the accuracy of the estimates at this stage does not justify firm conclusions as to which is the least expensive process. The major point to be derived is that the cost difference between magnesia and Wellman-Lord scrubbing has narrowed to a point where the two processes are very competitive. In making a choice between the two processes, reliability, flexibility, experience, and site-specific factors will have important influences on the decision, as well as more than comparative economics.

Energy requirements of the Wellman-Lord process are shown in Table 9 and Figure 21 along with the citrate and magnesia processes. The magnesia process has the lowest energy consumption and the citrate process has the largest. The energy needs of the magnesia process are about 13% less than the Wellman-Lord system when producing sulfuric acid. As would be expected, the manufacture of sulfuric acid by either magnesia scrubbing or Wellman-Lord is less energy intensive than the two sulfur production options.

TABLE 9. FGD PROCESS ENERGY REQUIREMENTS

Process	Electricity		Steam		Oil, MBtu/hr	Coal, MBtu/hr	Natural gas, MBtu/hr	Heat credit, MBtu/hr	Total equivalent energy consumption,[a] % of input energy
	MBtu/hr	kW	Reheat, MBtu/hr	Process, MBtu/hr					
Magnesia	67.2	7,468	71.3	0.6	117.7	-	-	(11.9)	5.6
Wellman Lord - H_2SO_4	64.4	7,269	61.2	156.4	-	-	-	(6.9)	6.5
Wellman Lord - Allied Chemical	62.0	6,890	61.2	164.7	12.0	10.0	-	(3.1)	7.4
Citrate	33.4	8,789	69.9	76.9	-	-	150.0	-	7.7

[a] Based on a 500-MW boiler efficiency of 90% for generation of steam and a gross heat rate of 9,000 Btu/kWh for generation of electricity.

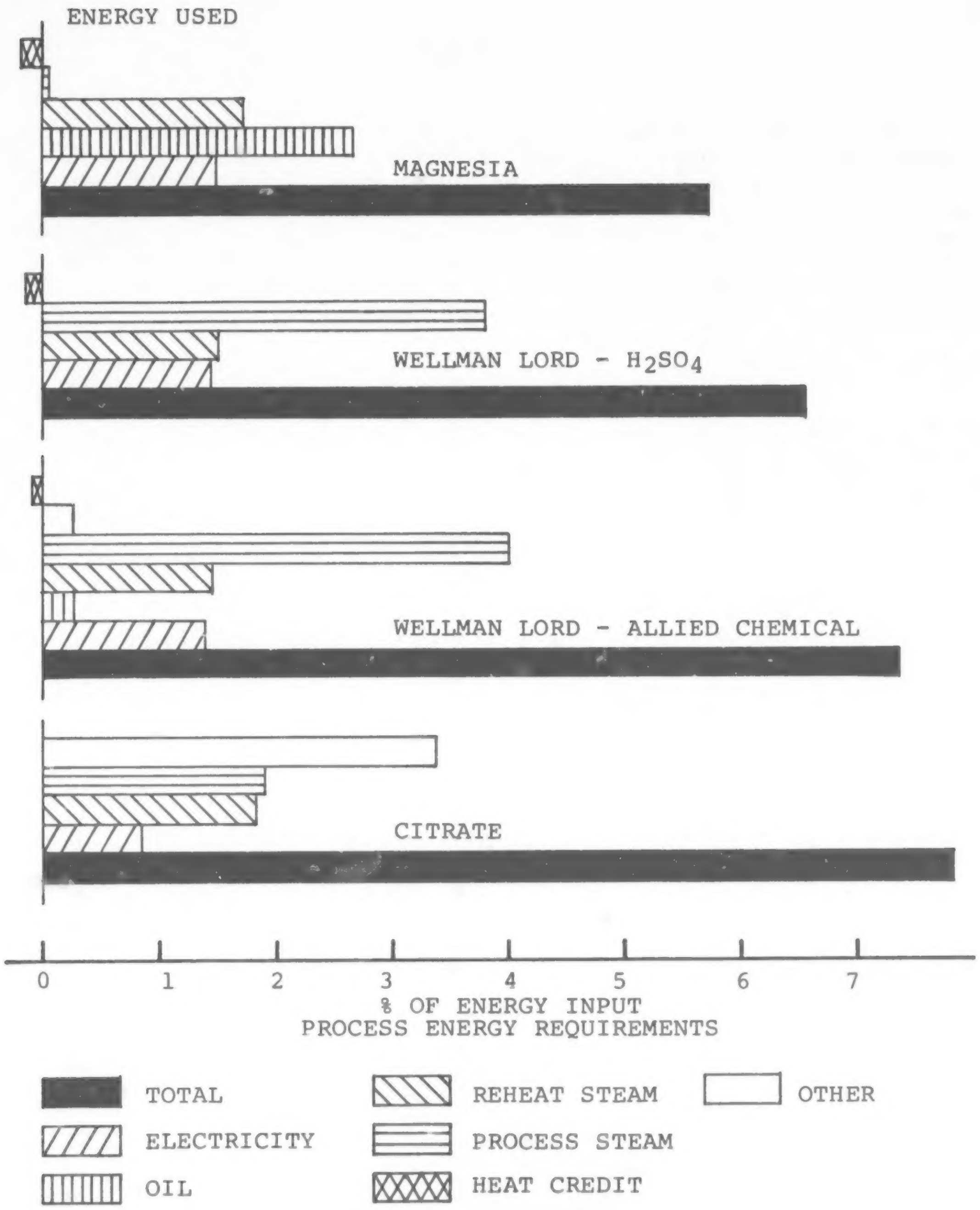

Figure 21. Process energy requirement for recovery processes.

EPILOGUE

In this paper, the key results from three on-going economic evaluation studies have been given. They are but a fraction of the results that will be presented in the final reports which will be issued at the end of the projects. The complete reports will include many case variations and sensitivities which will more fully describe the potential of each coal-cleaning and FGD process. Because all three projects are still underway, some results presented in this paper may be further refined but no major adjustments should be necessary.

APPENDIX A

PREMISES

COAL-CLEANING DESIGN AND ECONOMIC PREMISES

The design and economic premises for coal-cleaning plants follow in most respects the assumptions and procedures developed for FGD premises described in the following section. There are, however, some differences which must be recognized in making direct comparisons. These differences are described below.

Design Premises

Power Plant. The base-case conditions for coal-cleaning evaluations are a new 2000-MW midwestern power plant with a design heat rate of 9500 Btu/kWh operating at full capacity for 5500 hr/yr. The power plant life is assumed to be 30 years.

Coal Compositions. The full study uses coals containing 0.7, 2.0, 3.5, and 5.0% sulfur representing compositions that are typical of published information for over 350 coals with sulfur contents close to these levels. The 5.0% sulfur coal composition used in the evaluation discussed here is shown below.

Coal composition (5.0% sulfur, dry basis)	Wt % as received
Total sulfur	4.82
Pyritic sulfur	3.23
Sulfur as sulfate	0.06
Organic sulfur	1.53
Ash	16.1
Water*	3.5

*Air-dried moisture. Appropriate surface moistures are added depending on coal sizes.

Coal-Cleaning Plant. The coal-cleaning plants are assumed to be located at the power plant and are sized to supply the power plant coal demand. The PCC plants are based on a 90% Btu recovery and 6000 hr/yr of operation. The CCC plants are based on conversion and loss data supplied by the developers and 8000 hr/yr of operation. The cleaning plants have a 15-day raw coal and a 15-day clean coal storage based on power plant usage. The location, design, and size premises of waste ponds where required are identical to the FGD pond premises. PCC plants will have landfill disposal of solid wastes with mechanical compaction and an earth cover. The disposal site is located 1 mile from the coal preparation site.

Economic Premises

Other than an advanced project schedule and revised cost indexes the economic premises used for coal cleaning are the same as those used for FGD, as described in the following section. Costs are, of course, based on the coal-cleaning plant size and operating schedule.

Project Schedule. The coal-cleaning projects are assumed to begin in mid-1979 and end in mid-1982 with an average capital investment cost basis of the end of 1980. Annual revenue requirements are based on end of 1982 costs.

Direct Investment. Chemical Engineering cost indexes through 1977 and TVA projections of these indexes through 1983 are used to determine direct investments. The cost indexes and projections are shown below.

FGD Comparative Case

The FGD system used for comparison with the coal-cleaning processes is limestone scrubbing with 25% scrubber redundancy, 85% SO_x removal, and pond sludge disposal at the power plant and coal conditions used in the coal-cleaning premises. Capital and operating costs are also based on the coal-cleaning premises as described above.

FLUE GAS DESULFURIZATION DESIGN AND ECONOMIC PREMISES

The premises used for the ground-to-ground economic study and the Wellman-Lord reevaluation are discussed on the following page.

Coal Cleaning Cost Indexes and Projections

Year	1974	1975	1976	1977	1978[a]	1979[a]	1980[a]	1981[a]	1982[a]	1983[a]
Plant	165.4	182.4	192.1	204.1	221.4	240.2	259.4	278.9	299.8	322.3
Material[b]	171.2	194.7	205.8	220.9	240.8	262.5	286.1	309.0	333.7	360.4
Labor[c]	163.3	168.6	174.2	178.2	194.2	209.7	226.5	244.6	264.2	285.3

a. TVA projections.
b. Same as index in Chemical Engineering for "equipment, machinery, supports."
c. Same as index in Chemical Engineering for "construction labor."

Design

Base Case. The base case for conceptual design and preliminary cost estimating of FGD systems is a new 500-MW Midwestern power unit with a heat rate of 9000 Btu/kWh. The unit burns 3.5% sulfur coal (dry basis) with an as-fired heating value (HHV) of 10,500 Btu/lb and an ash content of 16%. The as-fired coal composition and flow rate for the base case design is shown below.

Coal composition (3.5% sulfur, dry basis)	Wt %, as fired	Lb/hr
Carbon	57.56	246,800
Hydrogen	4.14	17,700
Nitrogen	1.29	5,500
Oxygen	7.00	30,000
Sulfur	3.12	13,400
Chloride	0.15	600
Ash	16.00	68,600
Water	10.74	46,000
Total	100.00	428,600

Operating Life. The projected operating life of a new coal-fueled power unit is assumed to be 30 years representing a total of 127,500 hours of operation during the life of the plant. Operation during the first year is assumed to be 7000 hours.

Flue Gas Composition. Flue gas composition is based on the combustion of pulverized coal assuming a total air rate of the air preheater equivalent to 133% of the stoichiometric requirement. This includes 20% excess air to the boiler and 13% air inleakage at the air preheater. A horizontal, frontal-fired, coal-burning unit is assumed. It is assumed that 80% of the ash present in the coal is emitted as fly ash and 95% of the sulfur in the coal is emitted as SO_x. One percent of the sulfur emitted as SO_x is assumed to be SO_3 and the remainder SO_2. Flue gas rate and composition is tabulated below.

Flue Gas Composition and Properties

Component	Vol, %	Lb/hr
N_2	73.76	3,450,000
O_2	4.83	258,200
CO_2	12.31	904,200
SO_2	0.24	25,130
SO_3	0.0024	317
NO_x	0.06	3,009
HCl	0.01	661
H_2	8.79	264,500
		4,906,000 lb/hr (approx)
		1,543,000 aft^3/min at 300°F (approx)

Fly ash loading, gr/sft^3 (60°F) dry basis	6.65
Fly ash loading, gr/sft^3 (60°F) wet basis	6.06

Degree of SO_2 Removal. For the processes presented here, SO_2 removal is based on meeting the current SO_2 emission regulation of 1.2 lb SO_2 allowable emission/MBtu (M = one million) heat input.

Redundancy. No special redundancy is provided except spare pumps. The design does not include a bypass around either the ESP or the FGD units.

Reheat. Indirect steam reheat is used for all cases. Entrainment is estimated as 0.1% of the wet gas flow rate at the scrubber outlet for calculating the steam required for reheat.

Waste and Byproduct Management. An onsite disposal pond lined with impervious clay is used to contain the sulfite sludge from the limestone and lime processes. The pond is assumed to be located one mile from the scrubbing site. Thirty-day storage of byproduct sulfuric acid or sulfur is provided in the other processes.

Project Schedule. Projects are assumed to begin in mid-1977 and end in mid-1980, with an average capital investment cost basis of mid-1979. Direct investments are prepared using the average annual *Chemical Engineering* cost indexes and the TVA projections shown below. Although actual cost indexes are available for 1976-1978, TVA continues to use its projections for these years so that consistency with past estimates is maintained.

FGD Cost Indexes and Projections

Year	1973	1974	1975	1976[a]	1977[a]	1978[a]	1979[a]	1980[a]	1981[a]
Plant	144.1	165.4	182.4	197.9	214.7	232.9	251.5	271.6	293.3
Material[b]	141.9	171.2	194.7	210.3	227.1	245.3	264.9	286.1	309.0
Labor[c]	157.9	163.3	168.6	183.8	200.3	218.3	237.1	259.3	282.6

a. Projections.
b. Same as index in *Chemical Engineering* for "equipment, machinery, supports."
c. Same as index in *Chemical Engineering* for "construction labor."

Direct Investment Basis. Direct costs consist of materials and labor for equipment and installation, services and utilities, and pond construction. Services, utilities, and miscellaneous costs are estimated as 6% of the process areas subtotal. This covers such items as maintenance shops, stores, communications, railroad, and fire and service water facilities.

Indirect Investment Basis. Indirect costs consist of in-house engineering design and supervision, architect and engineering contractor expenses, contractor fees, and construction expenses. Construction facilities are considered a part of construction expenses. Consultant fees are not included. The engineering design and supervision, and the contingency factors are based on demonstration-level technology and experience. Indirect investment costs are estimated from the number of drawings required, man-hours of supervision and construction, and other factors related to the complexity of the process.

Allowances. Allowances are included for startup and modification, interest during construction, and working capital. Startup and modification allowances are estimated as 10% of total fixed investment for the recovery processes and 10% of the total fixed investment minus pond construction cost for the processes requiring a waste pond. Interest during construction is estimated as 12% of the subtotal fixed investment for each process. This factor is equivalent to the simple interest which would be accumulated at a 10%/yr rate assuming an incremental capital structure of 60% debt, 40% equity, and a 3-year project expenditure schedule as indicated below.

Project Expenditure Schedule

	Year			
	1	2	3	Total
Fraction of total expenditure as borrowed funds	0.15	0.30	0.15	0.60
Simple interest at 10%/yr as percent of total expenditure				
Year 1 debt	1.5	1.5	1.5	4.5
Year 2 debt	-	3.0	3.0	6.0
Year 3 debt	-	-	1.5	1.5
Accumulated interest as percent of total expenditure	1.5	4.5	6.0	12.0

Working Capital. Working capital consists of the total amount of money invested in raw materials and supplies carried in stock, finished products in stock, and semifinished products in the process of being manufactured; accounts receivable; cash kept on hand for payment of operating expenses such as salaries, wages, and raw material purchases; accounts payable; and taxes payable. For these premises, working capital is defined as the equivalent cost of 3 weeks of raw material costs, 7 weeks of direct costs, and 7 weeks of overhead costs.

Revenue Requirements

Direct Costs. Annual revenue requirements are based on 7000 hours of operation per year. Process operation schedules are assumed to be the same as the power plant operating profiles. Raw material, labor, and utility costs are projected to 1980. Maintenance costs are estimated on the basis of direct investment and are varied for each process according to the relative process complexity, and historical experience when available.

Indirect Costs. Following power industry practice, regulated company economics and the conventional method of considering the overall life of the power plant are used to establish capital charges. Straight-line depreciation of 3.3% is used.

Following Federal Energy Regulatory Commission (FERC) recommendations an interim replacements allowance factor is used in estimating annual revenue requirements to provide for the replacement of short-lived items. An average allowance of about 0.35% of the total investment is normally provided. However, to provide for the unknown life span of SO_x control facilities, a somewhat larger allowance factor of 0.7 is used. An insurance allowance of 0.5% of total depreciable capital investment is also included in the capital charges based on FERC practice. Property taxes are estimated as 1.5% of the total depreciable capital investment.

Cost of capital and income tax charges of 8.6% are applied to the unrecovered portion of capital investment, based on the debt-to-equity ratio of 60:40, bonds at 10% interest, and a 14% return on equity.

Overheads. Plant, administrative, and marketing overheads are costs which vary from company to company. With consideration of the various methods used in industry and illustrated in a variety of cost estimating sources, the following method for estimating overheads is used.

Plant overhead is estimated as 50% of the conversion costs excluding utilities. Administrative overhead is estimated as 10% of operating labor and supervision. Marketing byproducts is considered in the estimation of overheads as 10% of sales revenue.

Byproduct Sales. In estimating average annual revenue requirements, credit from sale of byproducts is deducted from the yearly projection of operating cost to obtain the net effect of the FGD process on the cost of power.

TABLE B-1. PCC PROCESS I

TOTAL CAPITAL INVESTMENT

(Dense-medium vessel, dense-medium cyclone, froth flotation)

Base case - 5% S coal

	Investment, $
Direct Investment	
Coal receiving and storage	8,841,000
Raw coal sizing	1,627,000
Coarse coal cleaning	1,585,000
Intermediate coal cleaning	2,249,000
Fine coal cleaning	2,696,000
Refuse disposal as landfill	3,058,000
Clean coal storage	8,261,000
Total areas	28,317,000
Services, utilities, and miscellaneous	1,699,000
Total direct investment	30,016,000
Indirect Investment	
Engineering design and supervision	2,521,000
Architect and engineering contractor	600,000
Construction expense	3,572,000
Contractor fees	1,009,000
Total indirect investment	7,702,000
Contingency	5,658,000
Total fixed investment	43,376,000
Other Capital Charges	
Allowance for startup and modifications	4,337,000
Interest during construction	6,073,000
Total depreciable investment	53,786,000
Land	3,686,000
Working capital	9,946,000
Total capital investment	67,418,000
Dollars of total capital per kW of generating capacity	33.71

Basis

Midwest location of coal-cleaning plant with project beginning mid-1979, ending mid-1982; average basis for cost scaling, end-1980; operating time, 6,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Fifteen-day raw coal and fifteen-day clean coal storage capacities (power plant basis).

Working capital provides for 3 weeks raw coal consumption, 7 weeks direct revenue costs (excluding Btu loss), and 7 weeks operating overheads.

Landfill site for refuse disposal located 1 mile from coal preparation plant.

TABLE B-2. PCC PROCESS I

ANNUAL REVENUE REQUIREMENTS

(Dense-medium vessel, dense-
medium cyclone, froth flotation)

Base case - 5% S coal	Annual quantity	Unit cost, $	Total annual cost, $
Direct Costs			
Raw materials			
Coal loss (Btu basis)	478,100 tons	31.58/ton	15,098,000
Total raw materials cost			15,098,000
Conversion costs			
Operating labor and supervision	144,000 man-hr	13.80/man-hr	1,987,000
Utilities			
Process water	45,300 kgal	0.13/kgal	6,000
Electricity	15,110,000 kWh	0.039/kWh	589,000
Diesel fuel	145,000 gal	0.70/gal	102,000
Process material magnetite, Grade E	2,760 tons	93.31/ton	257,000
Maintenance, 6% of direct investment			1,801,000
Analyses	4,000 man-hr	18.70/man-hr	75,000
Total conversion costs			4,817,000
Total direct costs			19,915,000
Indirect Costs			
Capital charges			
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			3,227,000
Average cost of capital and taxes at 8.6% of total capital investment			5,798,000
Overheads			
Plant, 50% of operating labor and supervision			993,000
Administrative, 10% of operating labor			199,000
Marketing, 10% of sales revenue			-
Total indirect costs			10,217,000
Gross annual revenue requirements			30,132,000
Byproduct Sales Revenue			
None			-
Total annual revenue requirements			30,132,000

	Mills/kWh	¢/lb sulfur removed
Equivalent unit revenue requirements	2.74	16.3

Basis

Midwest coal-cleaning plant location; time basis for scaling, mid-1982; plant life, 30 years; operating time, 6,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Total direct investment, $30,016,000; total depreciable investment, $53,786,000; and total capital investment, $67,418,000.

Raw coal (moisture-free): 4,840,000 tons/yr, 5% sulfur, 16.7% ash, 12,000 Btu/lb, and 4.17 lb S/MBtu.

Clean coal (moisture-free): 4,073,000 tons/yr, 3.67% sulfur, 10.09% ash, 13,000, and 2.84 lb S/MBtu.

TABLE B-3. PCC PROCESS II

TOTAL CAPITAL INVESTMENT

(Low-gravity D.M. cyclone, high-gravity D.M. cyclone, froth flotation)

Base case - 5% S coal

	Investment, $
Direct Investment	
Coal receiving and storage	8,841,000
Raw coal sizing	1,845,000
Low-gravity cleaning	3,564,000
High-gravity cleaning	1,782,000
Fine coal cleaning	4,706,000
Refuse disposal as landfill	3,058,000
Clean coal storage	6,397,000
Middling coal storage	4,632,000
Total areas	34,825,000
Services, utilities, and miscellaneous	2,090,000
Total direct investment	36,915,000
Indirect Investment	
Engineering design and supervision	3,101,000
Architect and engineering contractor	738,000
Construction expense	4,393,000
Contractor fees	1,240,000
Total indirect investment	9,472,000
Contingency	6,958,000
Total fixed investment	53,345,000
Other Capital Charges	
Allowance for startup and modifications	5,335,000
Interest during construction	7,468,000
Total depreciable investment	66,148,000
Land	3,703,000
Working capital	10,033,000
Total capital investment	79,884,000
Dollars of total capital per kW of generating capacity	39.94

Basis

- Midwest location of coal-cleaning plant with project beginning mid-1979, ending mid-1982; average basis for cost scaling, end-1980; operating time, 6,000 hr/yr.
- Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.
- Fifteen-day raw coal and fifteen-day clean coal storage capacities (power plant basis).
- Working capital provides for 3 weeks raw coal consumption, 7 weeks direct revenue costs (excluding Btu loss), and 7 weeks operating overheads.
- Landfill site for refuse disposal located 1 mile from coal preparation plant.

TABLE B-4. PCC PROCESS II

ANNUAL REVENUE REQUIREMENTS

(Low-gravity D.M. cyclone, high-gravity
D.M. cyclone, froth flotation)

Base case - 5% S coal

	Annual quantity	Unit cost, $	Total annual cost, $
Direct Costs			
Raw materials			
Coal loss (Btu basis)	458,650 tons	31.58/ton	14,484,000
Total raw materials cost			14,484,000
Conversion costs			
Operating labor and supervision	144,000 man-hr	13.80/man-hr	1,987,000
Utilities			
Process water	85,800 kgal	0.13/kgal	11,000
Electricity	27,384,000 kWh	0.039/kWh	1,068,000
Diesel fuel	148,000 [illegible]	0.70/gal	104,000
Process material, magnetite, Grade E	2,920 tons	93.31/ton	272,000
Maintenance, 6% of direct investment			2,215,000
Analyses	4,000 man-hr	18.70/man-hr	75,000
Total conversion costs			5,732,000
Total direct costs			20,216,000
Indirect Costs			
Capital charges			
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			3,969,000
Average cost of capital and taxes at 8.6% of total capital investment			6,870,000
Overheads			
Plant, 50% of operating labor and supervision			993,000
Administrative, 10% of operating labor			199,000
Marketing, 10% of sales revenue			-
Total indirect costs			12,031,000
Gross annual revenue requirements			32,247,000
Byproduct Sales Revenue			
None			
Total annual revenue requirements			32,247,000

	Mills/kWh	c/lb sulfur removed
Equivalent unit revenue requirements	2.93	16.3

Basis
- Midwest location of coal-cleaning plant; time basis for scaling, mid-1982; plant life, 30 years; operating time, 6,000 hr/yr.
- Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.
- Total direct investment, $36,915,000; total depreciable investment, $66,148,000; total capital investment, $79,384,000.
- Raw coal (moisture-free): 4,820,000 ton/yr, 5% sulfur, 16.7% ash, 12,000 Btu/lb, and 4.17 lb S/MBtu.
- Clean coal (moisture-free): 4,049,000 ton/yr, 3.51% sulfur, 9.25% ash, 13,100 Btu/lb, and 2.68 lb S/MBtu.

TABLE B-5. PCC PROCESS III

TOTAL CAPITAL INVESTMENT

(Dense-medium cyclone, concentrating table)

Base case - 5% S coal

	Investment, $
Direct Investment	
Coal receiving and storage	8,841,000
Raw coal sizing	2,438,000
Coarse coal cleaning	3,912,000
Fine coal cleaning	7,850,000
Refuse disposal as landfill	2,980,000
Clean coal storage	8,261,000
Total areas	34,282,000
Services, utilities, and miscellaneous	2,057,000
Total direct investment	36,339,000
Indirect Investment	
Engineering design and supervision	3,052,000
Architect and engineering contractor	727,000
Construction expense	4,324,000
Contractor fees	1,221,000
Total indirect investment	9,324,000
Contingency	6,849,000
Total fixed investment	52,512,000
Other Capital Charges	
Allowance for startup and modifications	5,251,000
Interest during construction	7,352,000
Total depreciable investment	65,115,000
Land	3,583,000
Working capital	10,007,000
Total capital investment	78,705,000
Dollars of total capital per kW of generating capacity	39.35

Basis

Midwest location of coal-cleaning plant with project beginning mid-1979, ending mid-1982; average basis for cost scaling, end-1980; operating time, 6,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Fifteen-day raw coal and fifteen-day clean coal storage capacities (power plant basis).

Working capital provides for 3 weeks raw coal consumpti , 7 weeks direct revenue costs (excluding Btu loss), and 7 weeks operating overheads.

Landfill site for refuse disposal located 1 mile from coal preparation plant.

TABLE B-6. PCC PROCESS III

ANNUAL REVENUE REQUIREMENTS

(Dense-medium cyclone, concentrating table)

Base case - 5% S coal

	Annual quantity	Unit cost, $	Total annual cost, $
Direct Costs			
Raw materials			
Coal loss (Btu basis)	471,800 tons	31.58/ton	14,889,000
Total raw materials cost			14,889,000
Conversion costs			
Operating labor and supervision	144,000 man-hr	13.80/man-hr	1,987,000
Utilities			
Process water	25,600 kgal	0.13/kgal	3,000
Electricity	13,459,000 kWh	0.039/kWh	525,000
Diesel fuel	138,000 gal	0.70/gal	97,000
Process material: magnetite, Grade E	1,970 tons	93.31/ton	184,000
Maintenance, 6% of direct investment			2,180,000
Analyses	4,000 man-hr	18.70/man-hr	75,000
Total conversion costs			5,051,000
Total direct costs			19,940,000
Indirect Costs			
Capital charges			
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			3,907,000
Average cost of capital and taxes at 8.6% of total capital investment			6,769,000
Overheads			
Plant, 50% of operating labor and supervision			993,000
Administrative, 10% of operating labor			199,000
Marketing, 10% of sales revenue			-
Total indirect costs			11,868,000
Gross annual revenue requirements			31,808,000
Byproduct Sales Revenue			
None			-
Total annual revenue requirements			31,808,000

	Mills/kWh	¢/lb sulfur removed
Equivalent unit revenue requirements	2.89	18.2

Basis

Midwest location of coal-cleaning plant; time basis for scaling, mid-1982; plant life, 30 years; operating time, 6,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Total direct investment, $9,324,000; total depreciable investment, $65,115,000; total capital investment, $78,705,000.

Raw coal (moisture-free): 4,855,000 ton/yr, 5% sulfur, 16.7% ash, 12,000 Btu/lb, and 4.17 lb S/MBtu.

Clean coal (moisture-free): 4,111,000 ton/yr, 3.78% sulfur, 10.60% ash, 12,800 Btu/lb, and 2.94 lb S/MBtu.

TABLE B- 7 . KVB PROCESS

TOTAL CAPITAL INVESTMENT

Base case -5% S coal

	Investment, $
Direct Investment	
Raw material handling and preparation	10,197,600
Sulfur oxidation	5,984,700
Reactor off-gas cleaning	10,889,300
Fine coal leaching	7,426,700
Coarse coal leaching	6,624,800
Product agglomeration and handling	11,328,800
Leach solution neutralization and water handling	5,913,200
Settling pond	16,203,000
Subtotal	74,568,100
Services, utilities, and miscellaneous	4,474,100
Total direct investment	79,042,200
Indirect Investment	
Engineering design and supervision	6,639,900
Architect and engineering contractor	1,586,300
Construction expense	9,407,800
Contractor fees	2,658,500
Total indirect investment	20,292,500
Contingency	19,866,900
Total fixed investment	119,201,600
Other Capital Charges	
Allowance for startup and modifications	11,920,200
Interest during construction	16,688,200
Total depreciable investment	147,810,000
Land	3,611,000
Working capital	19,945,200
Total capital investment	171,366,200
Dollars of total capital per kW of generating capacity	85.7

Basis

Midwest location of coal-cleaning plant with project beginning mid-1979, ending mid-1982; average basis for cost scaling, end-1980; operating time, 8,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Fifteen-day raw coal and fifteen-day clean coal storage capacities (power plant basis).

Working capital provides for 3 weeks raw coal consumption, 7 weeks direct revenue costs, and 7 weeks operating overheads.

Pond site for sludge disposal located 1 mile from coal preparation plant.

TABLE B-8. KVB PROCESS

ANNUAL REVENUE REQUIREMENTS

Base case - 5% S coal

	Annual quantity	Unit cost, $	Total annual cost, $
Direct Costs			
Raw materials			
Lime	197,603 tons	43.31/ton	8,558,200
Oxygen	297,600 tons	21.13/ton	6,288,300
NO_2	952 tons	665.28/ton	633,300
NaOH (50%)	152,880 tons	99.57/ton	15,222,300
Sodium lignin sulfonate	75,200 tons	83.17/ton	6,254,400
Natural gas	24,000 kft^3	2.93/kft^3	70,300
Total raw materials cost			37,026,800
Conversion costs			
Operating labor and supervision	152,000 man-hr	13.80/man-hr	2,097,600
Utilities			
Steam	5,349,838 MBtu	2.54/MBtu	13,588,600
Process water	2,663,074 kgal	0.09/kgal	239,700
Electricity	222,739,157 kWh	0.039/kWh	8,686,800
Maintenance, 6% of direct investment			4,742,500
Analysis	24,000 man-hr	18.70/man-hr	448,800
Total conversion costs			29,804,000
Total direct costs			66,830,800
Indirect Costs			
Capital charges			
Depreciation, interim replacements, and insurance at 6% of total depreciable investment			8,868,600
Average cost of capital and taxes at 8.6% of total capital investment			14,737,500
Overheads			
Plant, 50% of operating labor and supervision			1,048,800
Administrative, 10% of operating labor and supervision			209,800
Marketing, 10% of sales revenue			-
Total indirect costs			24,864,700
Gross annual revenue requirements			91,695,500
Byproduct Sales Revenue			
None			-
Total annual revenue requirements			91,695,500

	Mills/kWh	¢/lb sulfur removed
Equivalent unit revenue requirements	8.3	26.3

Basis

Midwest coal-cleaning plant location; time basis for scaling, mid-1982; plant life, 30 years; operating time, 8,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Total direct investment, $79,042,200; total depreciable investment, $147,810,000; and total capital investment, $171,366,200.

Raw coal (moisture-free): 4,578,400 tons/yr, 5.0% sulfur 16.7% ash, 12,000 Btu/lb, and 4.2 lb S/MBtu.

Clean coal (moisture-free): 4,336,000 tons/yr, 1.3% sulfur, 11.4% ash, 12,600 Btu/lb, and 1.0 lb S/MBtu.

TABLE B-9. TRW-"GRAVICHEM" PROCESS

TOTAL CAPITAL INVESTMENT

Base case - 5% S coal	Investment, $
Direct Investment	
Raw material handling and preparation	7,874,600
"Gravichem" separation	7,915,700
Float coal washing	7,201,100
Reactor - regenerator	20,539,000
Acetone leaching	12,553,100
Acetone recovery and coal drying	28,202,800
Leach solution concentration	3,104,400
Neutralization and pond water handling	1,792,300
Product agglomeration and handling	12,188,600
Utility water handling	1,065,200
Settling pond	8,219,500
Subtotal	110,656,300
Services, utilities, and miscellaneous	6,639,400
Total direct investment	117,295,700
Indirect Investment	
Engineering design and supervision	5,738,500
Architect and engineering contractor	1,387,900
Construction expense	13,028,600
Contractor fees	3,588,600
Total indirect investment	23,743,600
Contingency	28,207,900
Total fixed investment	169,247,200
Other Capital Charges	
Allowance for startup and modifications	16,924,700
Interest during construction	23,694,600
Total depreciable investment	209,866,500
Land	1,988,200
Working capital	16,194,400
Total capital investment	228,049,100
Dollars of total capital per kW of generating capacity	114.0

Basis

Midwest location of coal-cleaning plant with project beginning mid-1979, ending mid-1982; average basis for cost scaling, end-1980; operating time, 8,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Fifteen-day raw coal and fifteen-day clean coal storage capacities (power plant basis).

Working capital provides for 3 weeks raw coal consumption, 7 weeks direct revenue costs, and 7 weeks operating overheads.

Pond site for sludge disposal located 1 mile from coal preparation plant.

TABLE B-10. TRW-"GRAVICHEM" PROCESS

ANNUAL REVENUE REQUIREMENTS

Base case - 5% S coal

	Annual quantity	Unit cost, $	Total annual cost, $
Direct Costs			
Raw materials			
Lime	119,200 tons	43.31/ton	5,162,600
Oxygen	56,000 tons	21.13/ton	1,183,500
Acetone	2,872 tons	471.24/ton	1,353,400
Copperas	28,000 tons	72.07/ton	2,018,000
Sulfuric acid	83,200 tons	45.18/ton	3,759,000
Total raw materials cost			13,476,500
Conversion costs			
Operating labor and supervision	160,000 man-hr	13.80/man-hr	2,208,000
Utilities			
Steam	6,728,550 MBtu	2.54/MBtu	17,090,500
Process water	14,512,600 kgal	0.07/kgal	1,015,900
Electricity	182,028,933 kWh	0.039/kWh	7,099,100
Maintenance, 6% of direct investment			7,037,700
Analysis	32,000 man-hr	18.70/man-hr	598,400
Total conversion costs			35,049,600
Total direct costs			48,526,100
Indirect Costs			
Capital charges			
Depreciation, interim replacement, and insurance at 6% of total depreciable investment			12,592,000
Average cost of capital and taxes at 8.6% of total capital investment			19,612,200
Overheads			
Plant, 50% of operating labor and supervision			1,104,000
Administrative, 10% of operating labor and supervision			220,800
Marketing, 10% of sales revenue			123,500
Total indirect costs			33,652,500
Gross annual revenue requirements			82,178,600
Byproduct Sales Revenue			
Sulfur	23,300 long tons	53.00/long ton	(1,234,900)
Total annual revenue requirements			80,943,700

	Mills/kWh	¢/lb sulfur removed
Equivalent unit revenue requirements	7.4	27.4

Basis

Midwest coal-cleaning plant location; time basis for scaling, mid-1982; plant life, 30 years; operating time, 8,000 hr/yr.

Clean coal production capacity for 2,000-MW, coal-fired power plant operating at 9,550 Btu/kWh and 5,500 hr/yr.

Total direct investment, $117,295,700; total depreciable investment, $209,866,500; and total capital investment, $228,049,100.

Raw coal (moisture-free): 4,578,400 tons/yr, 5.0% sulfur, 16.7% ash, 12,000 Btu/lb, and 4.2 lb S/MBtu.

Clean coal (mositure-free): 4,364,800 tons/yr, 1.86% sulfur, 13.6% ash, 12,300 Btu/lb, and 1.5 lb S/MBtu.

TABLE B-11. KENNECOTT PROCESS

TOTAL CAPITAL INVESTMENT

Base case - 5% S coal	Investment, $
Direct Investment	
Raw materials handling and preparation	13,856,900
Reactor area	48,820,200
Coal filtration area	24,489,000
Product agglomeration and handling	28,343,900
Neutralization and water handling	6,151,800
Settling pond	13,961,900
Subtotal	135,623,700
Services, utilities, and miscellaneous	8,137,400
Total direct investment	143,761,100
Indirect Investment	
Engineering design and supervision	3,777,600
Architect and engineering contractor	877,700
Construction expense	15,321,300
Contractor fees	4,188,700
Total indirect investment	24,165,300
Contingency	33,585,300
Total fixed investment	201,511,700
Other Capital Charges	
Allowance for startup and modifications	20,151,200
Interest during construction	28,211,600
Total depreciable investment	249,874,500
Land	3,152,600
Working capital	29,188,800
Total capital investment	281,215,900
Dollars of total capital per kW of generating capacity	140.6

Basis

Midwest location of coal-cleaning plant with project beginning mid-1979, ending mid-1982; average basis for cost scaling, end-1980; operating time 8,000 hr/yr.

Clean coal production capacity for 2,000-MW, coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Fifteen-day raw coal and fifteen-day clean coal storage capacities (power plant basis).

Working capital provides for 3 weeks raw coal consumption, 7 weeks direct revenue costs, and 7 weeks operating overheads.

Pond site for sludge disposal located 1 mile from coal preparation plant.

TABLE B-12. KENNECOTT PROCESS

ANNUAL REVENUE REQUIREMENTS

Base case - 5% S coal

	Annual quantity	Unit cost, $	Total annual cost, $
Direct Costs			
Raw materials			
Lime	290,480 tons	43.31/ton	12,580,700
Oxygen	1,034,400 tons	21.13/ton	21,856,900
Sodium lignin sulfonate	171,200 tons	83.17/ton	14,238,700
Total raw materials cost			48,676,300
Conversion costs			
Operating labor and supervision	168,000 man-hr	13.80/man-hr	2,318,400
Process Btu loss	2,005,900 MBtu	1.36/MBtu	2,728,000
Steam	12,458,440 MBtu	2.54/MBtu	31,644,400
Process water	8,741,630 kgal	0.07/kgal	611,900
Electricity	660,230,321 kWh	0.039/kWh	25,749,000
Maintenance, 6% of direct investment			8,625,700
Analysis	32,000 man-hr	18.70/man-hr	598,400
Total conversion costs			72,275,800
Total direct costs			120,952,100
Indirect Costs			
Capital charges			
Depreciation, interim replacement, and insurance at 6% of total depreciable investment			14,992,500
Average cost of capital and taxes at 8.6% of total capital investment			24,184,600
Overheads			
Plant, 50% of operating labor and supervision			1,159,200
Administrative, 10% of operating labor and supervision			231,800
Marketing, 10% of sales revenue			-
Total indirect costs			40,568,100
Gross annual revenue requirements			161,520,200
Byproduct Sales Revenue			
None			-
Total annual revenue requirements			161,520,200

	Mills/kWh	¢/lb sulfur removed
Equivalent unit revenue requirements	14.7	48.9

Basis

Midwest coal-cleaning plant location; time basis for scaling, mid-1982; plant life, 30 years; operating time, 8,000 hr/yr.

Clean coal production capacity for 2,000-MW coal-fired power plant operating at 9,500 Btu/kWh and 5,500 hr/yr.

Total direct investment, $143,761,100; total depreciable investment, $249,874,500; and total capital investment, $281,215,900.

Raw coal (moisture-free): 5,249,600 tons/yr, 5.0% sulfur, 16.7% ash, 12,000 Btu/lb, and 4.2 lb S/MBtu.

Clean coal (moisture-free): 5,402,400 tons/yr, 1.8% sulfur, 13.6% ash, 12,300 Btu/lb, and 1.5 lb S/MBtu.

TABLE B-13. LIMESTONE SLURRY PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (hoppers, feeders, conveyors, elevators, bins, shaker and puller)	1,863,000	7.1
Feed preparation (feeders, crushers, ball mills, hoist, tanks, agitators, and pumps)	1,651,000	6.3
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,318,000	16.6
SO2 absorption (four TCA scrubbers including presaturators and entrainment separators, recirculation tanks, agitators, and pumps)	8,974,000	34.4
Stack gas reheat (four indirect steam reheaters)	1,282,000	4.9
Solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, and pond water return pumps)	1,680,000	6.4
Subtotal	19,768,000	75.7
Services, utilities, and miscellaneous	1,186,000	4.6
Total process areas excluding pond construction	20,954,000	80.3
Pond construction	5,145,000	19.7
Total direct investment	26,099,000	100.0
Indirect Investment		
Engineering design and supervision	1,218,000	4.7
Architect and engineering contractor	270,000	1.0
Construction expense	3,630,000	13.9
Contractor fees	1,145,000	4.4
Total indirect investment	6,263,000	24.0
Contingency	6,473,000	24.8
Total fixed investment	38,835,000	148.8
Other Capital Charges		
Allowance for startup and modifications	3,369,000	12.9
Interest during construction	4,660,000	17.9
Total depreciable investment	46,864,000	179.6
Land	1,030,000	3.9
Working capital	1,054,000	4.0
Total capital investment	48,948,000	187.5

Basis

Evaluation represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.

Stack gas reheat to 175°F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Disposal pond located 1 mile from power plant.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE B-14. LIMESTONE SLURRY PROCESS

SUMMARY OF AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of average annual revenue requirements
Direct Costs				
Raw materials				
Limestone	159,300 tons	7.00/ton	1,115,100	7.76
Total raw materials cost			1,115,100	7.76
Conversion costs				
Operating labor and supervision	25,990 man-hr	12.50/man-hr	324,900	2.26
Utilities				
Steam	489,800 MBtu	2.00/MBtu	979,600	6.81
Process water	243,400 kgal	0.12/kgal	29,200	0.20
Electricity	53,588,000 kWh	0.029/kWh	1,554,100	10.81
Maintenance				
Labor and material			2,040,200	14.19
Analyses	3,760 man-hr	17.00/man-hr	63,900	0.45
Total conversion costs			4,991,900	34.72
Total direct costs			6,107,000	42.48
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			2,811,800	19.56
Average cost of capital and taxes at 8.6% of total capital investment			4,209,500	29.28
Overheads				
Plant, 50% of conversion costs less utilities			1,214,500	8.45
Administrative, 10% of operating labor			32,500	0.23
Total indirect costs			8,268,300	57.52
Total average annual revenue requirements			14,375,300	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements	4.11	9.58	0.46	411

Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 35,000 short tons/yr; solids disposal 184,200 tons/yr calcium solids including only hydrate water.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $26,099,000; total depreciable investment, $46,864,000; and total capital investment, $48,948,000.
All tons shown are 2,000 lb.

TABLE B-15. LIME SLURRY PROCESS WITH CALCINATION

SUMMARY OF ESTIMATED CAPITAL INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in fuel;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (conveyors, elevators, feeder, silo, and bins)	2,570,000	9.0
Lime calcination (feeders, crusher, ball mill, fans, bins, rotary kiln, waste heat boiler, and elevators)	3,654,000	12.8
Feed preparation (feeders, slakers, tanks, agitators, and pumps)	660,000	2.3
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,318,000	15.2
SO_2 absorption (four TCA scrubbers including presaturators and entrainment separators, recirculation tanks, agitators, and pumps)	8,504,000	29.9
Stack gas reheat (four indirect steam reheaters)	1,282,000	4.5
Solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, and pond water return pumps)	1,616,000	5.7
Subtotal	22,604,000	79.4
Services, utilities, and miscellaneous	1,356,000	4.8
Total process areas excluding pond construction	23,906,000	84.2
Pond construction	4,505,000	15.8
Total direct investment	28,465,000	100.0
Indirect Investment		
Engineering design and supervision	1,683,000	5.9
Architect and engineering contractor	389,000	1.4
Construction expense	3,944,000	13.8
Contractor fees	1,223,000	4.3
Total indirect investment	7,239,000	25.4
Contingency	7,141,000	25.1
Total fixed investment	42,845,000	150.5
Other Capital Charges		
Allowance for startup and modifications	3,834,000	13.5
Interest during construction	5,142,000	18.1
Total depreciable investment	51,821,000	182.1
Land	909,000	3.2
Working capital	1,129,000	4.0
Total capital investment	53,859,000	189.3

Basis
- Evaluation represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
- Stack gas reheat to 175°F by indirect steam reheat.
- Minimum in-process storage; only pumps are spared.
- Disposal pond located 1 mile from power plant.
- Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
- Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE B-16. LIME SLURRY PROCESS WITH CALCINATION

SUMMARY OF AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of average annual revenue requirements
Direct Costs				
Raw materials				
Limestone	129,400 tons	7.00/ton	905,800	5.83
Coal	19,630 tons	25.00/ton	490,800	3.16
Total raw materials cost			1,396,600	8.99
Conversion costs				
Operating labor and supervision	37,670 man-hr	12.50/man-hr	470,900	3.03
Utilities				
Steam	488,340 MBtu	2.00/MBtu	976,700	6.29
Process water	235,600 kgal	0.12/kgal	28,300	0.18
Electricity	51,286,000 kWh	0.029/kWh	1,487,300	9.58
Heat credit	25,100 MBtu	2.00/MBtu	(50,200)	(0.32)
Maintenance				
Labor and material			2,052,000	13.21
Analyses	4,700 man-hr	17.00/man-hr	79,900	0.51
Total conversion costs			5,044,900	32.48
Total direct costs			6,441,500	41.47
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			3,109,300	20.02
Average cost of capital and taxes at 8.6% of total capital investment			4,631,900	29.83
Overheads				
Plant, 50% of conversion costs less utilities			1,301,400	8.38
Administrative, 10% of operating labor			47,100	0.30
Total indirect costs			9,089,700	58.53
Total average annual revenue requirements			15,531,200	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements	4.44	10.35	0.49	444

Basis
- Midwest plant location, 1980 revenue requirements.
- Remaining life of power plant, 30 yr.
- Power unit on-stream time, 7,000 hr/yr.
- Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
- Stack gas reheat to 175°F.
- Sulfur removed, 35,000 short tons/yr; solids disposal 153,600 tons/yr calcium solids including only hydrate water.
- Investment and revenue requirement for removal and disposal of fly ash excluded.
- Total direct investment, $28,465,000; total depreciable investment, $51,821,000; and total capital investment, $53,859,000.
- All tons shown are 2,000 lb.

TABLE B-17. LIME SLURRY PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in fuel;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (conveyors, elevators, feeder, silo, and bins)	1,978,000	8.3
Feed preparation (feeders, slakers, tanks, agitators, and pumps)	660,000	2.8
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,318,000	18.0
SO_2 absorption (four TCA scrubbers including presaturators and entrainment separators, recirculation tanks, agitators, and pumps)	8,504,000	35.5
Stack gas reheat (four indirect steam reheaters)	1,282,000	5.3
Solids disposal (onsite disposal facilities including feed tank, agitator, slurry disposal pumps, and pond water return pumps)	1,616,000	6.7
Subtotal	18,358,000	76.6
Services, utilities, and miscellaneous	1,101,000	4.6
Total process areas excluding pond construction	19,459,000	81.2
Pond construction	4,505,000	18.8
Total direct investment	23,964,000	100.0
Indirect Investment		
Engineering design and supervision	1,095,000	4.6
Architect and engineering contractor	243,000	1.0
Construction expense	3,391,000	14.1
Contractor fees	1,073,000	4.5
Total indirect investment	5,802,000	24.2
Contingency	5,953,000	24.8
Total fixed investment	35,719,000	149.0
Other Capital Charges		
Allowance for startup and modifications	3,121,000	13.0
Interest during construction	4,286,000	17.9
Total depreciable investment	43,126,000	179.9
Land	895,000	3.7
Working capital	1,298,000	5.4
Total capital investment	45,319,000	189.0

Basis

Evaluation represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.

Stack gas reheat to 175°F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Disposal pond located 1 mile from power plant.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE B-18. LIME SLURRY PROCESS

SUMMARY OF AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; onsite solids disposal)

	Annual quantity	Unit cost, $	Total annual cost, $	% of average annual revenue requirements
Direct Costs				
Raw materials				
Lime	68,600 tons	42.00/ton	2,881,200	19.35
Total raw materials cost			2,881,200	19.35
Conversion costs				
Operating labor and supervision	25,990 man-hr	12.50/man-hr	324,900	2.18
Utilities				
Steam	489,900 MBtu	2.00/MBtu	979,600	6.58
Process water	232,600 kgal	0.12/kgal	27,900	0.19
Electricity	47,008,000 kWh	0.029/kWh	1,363,200	9.15
Maintenance				
Labor and material			1,691,900	11.36
Analyses	3,760 man-hr	17.00/man-hr	63,900	0.42
Total conversion costs			4,451,400	29.89
Total direct costs			7,332,600	49.24
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			2,587,600	17.38
Average cost of capital and taxes at 8.6% of total capital investment			3,897,400	26.17
Overheads				
Plant, 50% of conversion costs less utilities			1,040,400	6.99
Administrative, 10% of operating labor			32,500	0.22
Total indirect costs			7,557,900	50.76
Total average annual revenue requirements			14,890,500	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements	4.25	9.93	0.47	425

Basis
- Midwest plant location, 1980 revenue requirements.
- Remaining life of power plant, 30 yr.
- Power unit on-stream time, 7,000 hr/yr.
- Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
- Stack gas reheat to 175°F.
- Sulfur removed, 35,000 short tons/yr; solids disposal 153,600 tons/yr calcium solids including only hydrate water.
- Investment and revenue requirement for removal and disposal of fly ash excluded.
- Total direct investment, $23,964,000; total depreciable investment, $43,126,000; and total capital investment, $45,319,000.
- All tons shown are 2,000 lb.

TABLE B-19. MAGNESIA SLURRY - REGENERATION PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; 6.5 tons/hr 100% H_2SO_4)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (conveyors, silos, bins, and feeders)	1,031,000	2.8
Feed preparation (mixer, tank, agitator, and pump)	447,000	1.2
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,318,000	11.6
SO_2 absorption (four spray grid scrubbers including entrainment separators, tanks, agitators, and pumps)	5,874,000	15.8
Stack gas reheat (four indirect steam reheaters)	1,407,000	3.8
Chloride purge (four chloride scrubbers and entrainment separators, tanks, agitators, and pumps)	2,766,000	7.5
Slurry processing (centrifuges, conveyor, tank, agitator, and pumps)	1,499,000	4.0
Cake drying (dryer, conveyors, silos, fans, tank, and pumps)	5,747,000	15.5
Calcination (calciner, preheater, solids cooler, waste heat boiler, conveyors, silos, fans, and bins)	2,281,000	6.2
Acid production (complete contact unit for sulfuric acid production)	8,340,000	22.5
Acid storage (storage and shipping facilities for 30-day production of sulfuric acid)	1,265,000	3.4
Subtotal	34,975,000	94.3
Services, utilities, and miscellaneous	2,099,000	5.7
Total direct investment	37,074,000	100.0
Indirect Investment		
Engineering design and supervision	2,863,000	7.7
Architect and engineering contractor	528,000	1.4
Construction expense	5,015,000	13.5
Contractor fees	1,495,000	4.1
Total indirect investment	9,901,000	26.7
Contingency	9,395,000	25.3
Total fixed investment	56,370,000	152.0
Other Capital Charges		
Allowance for startup and modifications	5,637,000	15.2
Interest during construction	6,764,000	18.3
Total depreciable investment	68,771,000	185.5
Land	27,000	0.1
Working capital	1,495,000	4.0
Total capital investment	70,293,000	189.6

Basis

Evaluation represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.

Stack gas reheat to 175°F by indirect steam reheat.

Minimum in-process storage; only pumps are spared.

Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.

Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE B-20. MAGNESIA SLURRY - REGENERATION PROCESS

SUMMARY OF AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS

(500-MW new coal-fired power unit, 3.5% S in fuel;
1.2 lb SO_2/MBtu heat input allowable emission; 108,000 tons/yr 100% H_2SO_4)

	Annual quantity	Unit cost, $	Total annual cost, $	% of net average annual revenue requirements
Direct Costs				
Raw materials				
Magnesium oxide	1,470 tons	300.00/ton	441,000	2.41
Catalyst	1,800 liters	2.50/liter	4,500	0.02
Hydrated lime	2,800 tons	54.00/ton	151,200	0.83
Total raw materials cost			596,700	3.26
Conversion costs				
Operating labor and supervision	47,500 man-hr	12.50/man-hr	593,800	3.24
Utilities				
Fuel oil (No. 6)	5,720,000 gal	0.40/gal	2,288,000	12.49
Steam	503,400 MBtu	2.00/MBtu	1,006,800	5.49
Process water	2,359,200 kgal	0.12/kgal	283,100	1.54
Electricity	52,277,400 kWh	0.029/kWh	1,516,000	8.27
Heat credit	83,400 MBtu	2.00/MBtu	(166,800)	(0.91)
Maintenance				
Labor and material			2,595,200	14.16
Analyses	8,500 man-hr	17.00/man-hr	144,500	0.79
Total conversion costs			8,260,600	45.07
Total direct costs			8,857,300	48.33
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			4,126,300	22.52
Average cost of capital and taxes at 8.6% of total capital investment			6,045,200	32.99
Overheads				
Plant, 50% of conversion cost less utilities			1,666,800	9.10
Administrative, 10% of operating labor			59,400	0.32
Marketing, 10% of byproduct sales revenue			270,000	1.47
Total indirect costs			12,167,700	66.40
Gross average annual revenue requirements			21,025,000	114.73
Byproduct Sales Revenue				
100% sulfuric acid	108,000 tons	25.00/ton	(2,700,000)	(14.73)
Net average annual revenue requirements			18,325,000	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements (net)	5.24	12.22	0.58	524

Basis

Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 35,000 short tons/yr.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $37,074,000; total depreciable investment, $68,771,000; and total capital investment, $70,293,000.
All tons shown are 2,000 lb.

TABLE B-21. WELLMAN-LORD SCRUBBING/SULFURIC ACID PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; 14.3 tons/hr 100% H_2SO_4)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (conveyors, silos, feeders, bins, shaker, tanks, and pumps)	1,043,000	2.8
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack, gas ducts from chloride scrubber to absorber)	4,699,000	12.4
SO_2 absorption (four absorbers and entrainment separators, tanks, pumps, filters, agitators, and heat exchangers)	8,329,000	21.8
Stack gas reheat (four indirect steam reheaters)	1,179,000	3.1
Chloride purge (four chloride scrubbers and entrainment separators, tanks, agitators, and pumps)	2,644,000	7.0
Sulfate crystallization (evaporator-crystallizer, heat exchanger, pumps, agitator, tank, dryer, conveyors, centrifuge, bin, silo, and feeder)	1,907,000	5.0
SO_2 regeneration (evaporators, heat exchangers, strippers, tanks, agitators, pumps, blower, and condensers)	7,989,000	21.1
Acid production (complete contact unit for sulfuric acid production)	6,820,000	18.0
Acid storage (storage and shipping facilities for 30-day production of sulfuric acid)	1,163,000	3.1
Subtotal	35,773,000	94.3
Services, utilities, and miscellaneous	2,146,000	5.7
Total direct investment	37,919,000	100.0
Indirect Investment		
Engineering design and supervision	2,520,000	6.6
Architect and engineering contractor	630,000	1.7
Construction expense	5,110,000	13.5
Contractor fees	1,521,000	4.0
Total indirect investment	9,781,000	25.8
Contingency	9,540,000	25.2
Total fixed investment	57,240,000	151.0
Other Capital Charges		
Allowance for startup and modifications	5,724,000	15.1
Interest during construction	6,869,000	18.1
Total depreciable investment	69,833,000	184.2
Land	28,000	0.1
Working capital	1,587,000	4.1
Total capital investment	71,448,000	188.4

Basis
- Evaluation represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
- Stack gas reheat to 175°F by indirect steam reheat.
- Minimum in-process storage; only pumps are spared.
- Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
- Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE B-22. WELLMAN-LORD SCRUBBING/SULFURIC ACID PROCESS

SUMMARY OF AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; 100,300 tons/yr 100% H_2SO_4)

	Annual quantity	Unit cost, $	Total annual cost, $	% of average annual revenue requirements
Direct Costs				
Raw materials				
Sodium carbonate	6,860 tons	103.00/ton	706,600	3.62
Catalyst	2,000 liters	2.50/liter	5,000	0.03
Hydrated lime	2,800 tons	54.00/ton	151,200	0.77
Total raw materials cost			862,800	4.42
Conversion costs				
Operating labor and supervision	46,500 man-hr	12.50/man-hr	581,300	2.98
Utilities				
Steam	1,523,080 MBtu	2.00/MBtu	3,046,200	15.59
Process water	6,105,760 kgal	0.12/kgal	732,700	3.75
Electricity	50,882,670 kWh	0.029/kWh	1,475,600	7.56
Heat credit	48,360 MBtu	2.00/MBtu	(96,700)	(0.50)
Maintenance				
Labor and material			2,654,300	13.59
Analyses	8,500 man-hr	17.00/man-hr	144,500	0.74
Total conversion costs			8,537,900	43.71
Total direct costs			9,400,700	48.13
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			4,190,000	21.45
Average cost of capital and taxes at 8.6% of total capital investment			6,144,500	31.46
Overheads				
Plant, 50% of conversion cost less utilities			1,690,100	8.65
Administrative, 10% of operating labor			58,100	0.30
Marketing, 10% of byproduct sales revenue			269,400	1.11
Total indirect costs			12,352,100	62.97
Gross average annual revenue requirements			21,752,800	110.10
Byproduct Sales Revenue				
100% sulfuric acid	100,300 tons	25.00/ton 100% H_2SO_4	(2,507,500)	(10.27)
Sodium sulfate	8,110 tons	23.00/ton	(186,500)	(0.83)
Net average annual revenue requirements			19,058,800	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements (net)	5.44	12.71	0.61	545

Basis
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 yr.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 35,000 short tons/yr.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $37,919,000; total depreciable investment, $69,833,000; and total capital investment, $71,441,000.
All tons shown are 2,000 lb.

TABLE B-23. WELLMAN-LORD SCRUBBING/ALLIED CHEMICAL COAL/SO_2 REDUCTION PROCESS

SUMMARY OF ESTIMATED CAPITAL INVESTMENT

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; 4.7 tons/hr elemental S)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (conveyors, silos, feeders, bins, shaker, tanks, and pumps)	1,056,000	2.7
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack, gas ducts from chloride scrubber to absorber)	4,699,000	11.9
SO_2 absorption (four absorbers and entrainment separators, tanks, pumps, filters, agitators, heat exchangers)	8,344,000	21.2
Stack gas reheat (four indirect steam reheaters)	1,179,000	3.0
Chloride purge (four chloride scrubbers and entrainment separators, tank, agitators, and pumps)	2,644,000	6.7
Sulfate crystallization (evaporator crystallizer, heat exchanger, pumps, agitator, tank, dryer, conveyors, centrifuge, bin, silo, and feeder)	1,919,000	4.9
SO_2 regeneration (evaporators, heat exchangers, stripper, tanks, agitators, pumps, blower, and condensers)	8,064,000	20.5
Sulfur production (complete coal reduction unit)	8,400,000	21.3
Sulfur storage (storage and shipping facilities for 30-day production of sulfur)	593,000	1.5
Subtotal	36,898,000	93.7
Services, utilities, and miscellaneous	2,214,000	5.6
Total direct investment excluding pond construction	39,112,000	99.3
Pond construction	269,000	0.7
Total direct investment	39,381,000	100.0
Indirect Investment		
Engineering design and supervision	2,789,000	7.0
Architect and engineering contractor	692,000	1.8
Construction expense	5,286,000	13.4
Contractor fees	1,566,000	4.0
Total indirect investment	10,333,000	26.2
Contingency	9,620,000	24.4
Total fixed investment	59,334,000	150.7
Other Capital Charges		
Allowance for startup and modifications	5,907,000	15.0
Interest during construction	7,120,000	18.0
Total depreciable investment	72,361,000	183.7
Land	64,000	0.2
Working capital	1,765,000	4.5
Total capital investment	74,190,000	188.4

Basis

Midwest plant location represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Disposal pond located 1 mile from power plant.
Investment requirement for fly ash removal and disposal excluded; FGD process investment estimate begins with common plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE B-24. WELLMAN-LORD SCRUBBING/ALLIED CHEMICAL COAL/SO_2 REDUCTION PROCESS

SUMMARY OF AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; 32,690 tons/yr elemental S)

	Annual quantity	Unit cost, $	Total annual cost, $	% of net average annual revenue requirements
Direct Costs				
Raw materials				
Sodium carbonate	7,440 tons	103.00/ton	766,300	3.57
Coal	25,370 tons	26.50/ton	672,300	3.13
Sand	180 tons	7.50/ton	1,400	0.01
Catalyst			3,800	0.02
Hydrated lime	2,800 tons	54.00/ton	151,200	0.70
Total raw materials cost			1,595,000	7.43
Conversion costs				
Operating labor and supervision	46,500 man-hr	12.50/man-hr	581,300	2.71
Utilities				
Fuel oil (No. 6)	582,580 gal	0.40/gal	233,000	1.08
Steam	1,581,820 MBtu	2.00/MBtu	3,163,600	14.72
Process water	5,219,260 kgal	0.12/kgal	626,300	2.92
Electricity	48,230,700 kWh	0.029/kWh	1,398,700	6.51
Heat credit	21,640 MBtu	2.00/MBtu	(43,300)	(0.20)
Maintenance				
Labor and material			2,756,700	12.83
Analyses	8,800 man-hr	17.00/man-hr	149,600	0.70
Total conversion costs			8,865,900	41.27
Total direct costs			10,460,900	48.70
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			4,341,700	20.71
Average cost of capital and taxes at 8.6% of total capital investment			6,380,000	29.71
Overheads				
Plant, 50% of conversion cost less utilities			1,743,800	8.12
Administrative, 10% of operating labor			58,100	0.27
Marketing, 10% of byproduct sales revenue			167,400	0.78
Total indirect costs			12,691,000	59.09
Gross average annual revenue requirements			23,151,900	107.79
Byproduct Sales Revenue				
Sulfur	32,690 tons	45.00/ton S	(1,471,100)	(6.85)
Sodium sulfate	8,800 tons	23.00/ton	(202,400)	(0.94)
Net average revenue requirements			21,478,400	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements (net)	6.14	14.32	0.68	614

Basis
- Midwest plant location, 1980 revenue requirements.
- Remaining life of power plant, 30 yr.
- Power unit on-stream time, 7,000 hr/yr.
- Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
- Stack gas reheat to 175°F.
- Sulfur removed, 35,000 short tons/yr.
- Investment and revenue requirement for removal and disposal of fly ash excluded.
- Total direct investment, $39,381,000; total depreciable investment, $72,361,000; and total capital investment, $74,190,000.
- All tons shown are 2,000 lb.

TABLE B-25. CITRATE PROCESS

SUMMARY OF ESTIMATED FIXED INVESTMENT[a]

(500-MW new coal-fired power unit, 3.5% S in coal;
1.2 lb SO_2/MBtu heat input allowable emission; 4.8 tons/hr elemental S)

	Investment, $	% of total direct investment
Direct Investment		
Materials handling (conveyors and bins)	804,000	2.1
Feed preparation (conveyors, tanks, agitators, pumps, and feeders)	118,000	0.3
Gas handling (common feed plenum and booster fans, gas ducts and dampers from plenum to absorber, exhaust gas ducts and dampers from absorber to reheater and stack)	4,368,000	11.2
SO_2 absorption (four packed tower absorbers including presaturators and mist eliminators, surge tanks, centrifugal pumps, compressor, and strippers)	14,223,000	36.6
Stack gas reheat (four indirect steam reheaters)	1,294,000	3.3
Chloride purge (feeder, tank, agitator, and pump)	83,000	0.2
SO_2 reduction (reactor tanks, aging tanks, agitators, and centrifugal pumps)	1,303,000	3.4
Sulfur separation and removal (flotation tanks, rotary drum filter, pumps, slurry tank, heat exchanger, settling tank, heaters, and flash drum)	2,118,000	5.5
Sulfur storage and shipping (sulfur receiving pit, heaters, sulfur pump, and storage tank)	814,000	2.1
Sulfate removal (coolers, agitators, centrifuge, tanks, pumps, and refrigeration)	985,000	2.5
H_2S generation (battery limit plant)	5,850,000	15.1
H_2 generation (battery limit plant)	4,680,000	12.0
Subtotal	36,640,000	94.3
Services, utilities, and miscellaneous	2,198,000	5.7
Total direct investment	38,838,000	100.0
Indirect Investment		
Engineering design and supervision	3,273,000	8.4
Architect and engineering contractor	818,000	2.1
Construction expense	5,208,000	13.4
Contractor fees	1,548,000	4.0
Total indirect investment	10,847,000	27.9
Contingency	9,937,000	25.6
Total fixed investment	59,622,000	153.5
Other Capital Charges		
Allowance for startup and modifications	5,962,000	15.4
Interest during construction	7,155,000	18.4
Total depreciable investment	72,739,000	187.3
Land	39,000	0.1
Working capital	2,140,000	5.5
Total capital investment	74,918,000	192.9

[a] Basis
Evaluation represents project beginning mid-1977, ending mid-1980. Average cost basis for scaling, mid-1979.
Stack gas reheat to 175°F by indirect steam reheat.
Minimum in-process storage; only pumps are spared.
Investment requirements for fly ash removal and disposal excluded; FGD process investment estimate begins with common feed plenum downstream of the ESP.
Construction labor shortages with accompanying overtime pay incentive not considered.

TABLE 8-26. CITRATE PROCESS

TOTAL AVERAGE ANNUAL REVENUE REQUIREMENTS - REGULATED UTILITY ECONOMICS[a]

(500-MW new coal-fired power unit, 3.5% S in coal
1.2 lb SO_2/MBtu heat input allowable emission; 33,840 tons/yr elemental S)

	Annual quantity	Unit cost, $	Total annual cost, $	% of net average annual revenue requirements
Direct Costs				
Raw materials				
Lime	2,870 tons	42.00/ton	120,500	0.52
Soda ash	2,630 tons	103.00/ton	270,900	1.16
Citric acid	230 tons	1,340.00/ton	308,200	1.32
Natural gas	1,050,000 kft^3	3.50/kft^3	3,675,000	15.78
Catalyst			21,000	0.09
Total raw material cost			4,395,600	18.87
Conversion costs				
Operating labor and supervision	67,920 man-hr	12.50/man-hr	849,000	3.64
Utilities				
Steam	1,027,500 MBtu	2.00/MBtu	2,055,000	8.82
Process water	2,492,500 kgal	0.12/kgal	299,100	1.28
Electricity	68,530,000 kWh	0.029/kWh	1,987,400	8.53
Maintenance				
Labor and material			2,330,300	10.01
Analyses	10,600 man-hr	17.00/man-hr	180,200	0.77
Total conversion costs			7,701,000	33.05
Total direct costs			12,096,600	51.92
Indirect Costs				
Capital charges				
Depreciation, interim replacements, and insurance at 6.0% of total depreciable investment			4,364,300	18.73
Average cost of capital and taxes at 8.6% of total capital investment			6,442,900	27.67
Overheads				
Plant, 50% of conversion costs less utilities			1,679,800	7.21
Administrative, 10% of operating labor			84,900	0.36
Marketing, 10% of byproduct sales revenue			152,300	0.65
Total indirect costs			12,724,200	54.62
Gross average annual revenue requirements			24,820,800	106.54
Byproduct Sales Revenue				
Sulfur	33,840 tons	45.00/tons	(1,522,800)	(6.54)
Net average revenue requirements			23,298,000	100.00

	Mills/kWh	$/ton coal burned	$/MBtu heat input	$/ton S removed
Equivalent unit revenue requirements (net)	6.66	15.53	0.74	666

a. **Basis**
Midwest plant location, 1980 revenue requirements.
Remaining life of power plant, 30 years.
Power unit on-stream time, 7,000 hr/yr.
Coal burned, 1,500,100 tons/yr, 9,000 Btu/kWh.
Stack gas reheat to 175°F.
Sulfur removed, 35,000 short tons/yr.
Investment and revenue requirement for removal and disposal of fly ash excluded.
Total direct investment, $38,838,000; total depreciable investment, $72,739,000; and total capital investment, $74,918,000.
All tons shown are 2,000 pounds.

COMBINED COAL CLEANING AND FGD

James D. Kilgroe

U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Research Triangle Park, N. C.

ABSTRACT

Physical coal cleaning (PCC) can be used to attain moderate reductions in the ash and sulfur levels of U.S. coals. PCC can thus be used to generate compliance fuel for the less stringent State and Federal standards governing fossil fuel fired steam generators. The sulfur reduction requirements and emission levels which are likely to be specified in the revised New Source Performance Standards (NSPS) for electric utility boilers will preclude the use of coal cleaning as a sole method of complying with these flue gas desulfurization (FGD) regulations.

The combined use of physical coal cleaning and flue gas desulfurization (PCC + FGD) will be the most cost effective method of complying with emission regulations,if the reduction in FGD and non-FGD costs which result from using cleaned coal are greater than the costs of PCC. Reductions in FGD costs by PCC can result from a reduction in the volume of flue gas treated (partial scrubbing) or the amount of sulfur removed from the flue gas stream. Reductions in fuel sulfur variability by PCC can lower design safety margins needed to ensure compliance for all fuel sulfur values. Non-FGD cost benefits can result from reduced boiler operation and maintenance costs, reduced transportation costs, reduced ash disposal costs,and reduced coal pulverization costs.

Utility boilers which use high sulfur coals and which require sulfur removals less than 75 percent are likely candidates for PCC + FGD. If the revised NSPS for utility boilers require 90 percent sulfur removal and do not specify an emission floor, then PCC + FGD may not be competitive with FGD unless there are substantial non-FGD cost benefits associated with cleaning.

The range of applications for PCC + FGD in small non-base-loaded utility boilers and industrial boilers may be different from those cited for base-loaded utility boilers. The differentials between PCC and FGD costs for these smaller units may result in different optimal solutions for the range of alternative control strategies.

INTRODUCTION

The Clean Air Act Amendments of 1977 will have a substantial impact on the costs and technologies used to comply with State and Federal SO_2 emission regulations. Regulatory actions in response to these amendments will result in tightening of emission standards in existing boilers in order to meet ambient air quality standards in non-attainment areas. Revised new source performance standards (NSPS) will be set for coal fired utility boilers and NSPS will be promulgated for industrial boilers.

Physical coal cleaning (PCC) can be used to attain moderate reductions in the ash and sulfur levels of U.S. coals. PCC can thus be used to generate compliance fuel for the less stringent state standards and current NSPS governing fossil fuel fired steam generators. The sulfur reduction requirements and emission levels which are likely to be specified for the revised NSPS for electric utility boilers will preclude the use of coal cleaning as a sole method of complying with these regulations.

Previous studies have shown that in some instances combinations of coal cleaning and flue gas desulfurization (FGD) can be a more cost effective emission control technique than FGD alone. This paper summarizes the potential of PCC as a method of coal desulfurization and evaluates conditions under which PCC can be used in conjunction with FGD to reduce the cost of complying with SO_2 emission regulations.

AIR POLLUTION REGULATIONS

The U.S. EPA is developing and implementing air pollution control regulations in accordance with the provisions of the Clean Air Act and its amendments. Stationary source emission standards are designed to regulate the quantities of pollutants emitted from point sources, whereas ambient air quality standards are designed to regulate the concentrations of pollutants in the atmosphere.

Ambient Air Quality Standards

Under Section 109 of the Clean Air Act of 1970, the U.S. EPA has established national primary and secondary ambient air quality standards (NAAQS) to protect human health and public welfare, respectively. State Implementation Plans (SIPs) which must be approved by the U.S. EPA are used to achieve permissible air quality levels for certain "criteria" pollutants including: total particulates, sulfur dioxide and nitrogen oxides.

The 1977 Clean Air Act Amendments require EPA to review the NAAQS no later than December 31, 1980 and at 5 year intervals thereafter. If warranted, revisions to the NAAQS are to be made. Subsequent to changes in the NAAQS, each state is required to modify its SIP to comply with the new air quality standards. Existing coal fired boilers are regulated under SIP regulations. Emission limits for these sources may range from about 0.2 lb $SO_2/10^6$ Btu* to above 6.0 lb $SO_2/10^6$ Btu depending on the boiler site.

New Source Performance Standards

New Source Performance Standards are issued by the U.S. EPA in accordance with Section 111 of the 1970 Clean Air Act. These standards of performance are applied to new and modified source categories designated by EPA. Many provisions of the Clean Air Act Amendments of 1977 are aimed specifically at coal firing sources, and the restrictions applied are much more rigorous than in the past. Where the original New Source Performance Standard for SO_2 emissions applying to large coal fired boilers permitted the emission of 1.2 lb SO_2/million Btu, the amended Act specified that the revised NSPS "...shall reflect the degree of emission limitation and the percentage reduction achievable through application of the best technological system of continuous emission reduction...;" i.e., a percentage reduction will be required in addition to the maintenance of emissions below an upper limit. Any cleaning of the fuel which reduces the pollution characteristics of the fuel after extraction and prior to combustion may be credited to the pollution percentage reduction requirement.

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* English to metric unit conversion factors are given at the end of this paper.

There is still considerable uncertainty concerning the revised NSPS for coal fired steam electric utility boilers which are to be promulgated by EPA (EPA, 1978a). At least four full and seven partial SO_2 control alternatives are currently under consideration (EPA, 1978b). The full control options generally specify emission limits of 0.55 to 0.80 lb $SO_2/10^6$ Btu with 90 to 95 percent sulfur control expressed on an annual basis. The partial control options generally include an emission limit, a percentage removal requirement, and a maximum control or emission floor below which the full sulfur reduction requirement does not apply. The partial control options typically result in reduced emission control costs where the emission floor can be met by burning low sulfur coals and partial scrubbing.

EPA is also considering NSPS for industrial boilers. Alternative standards now being studied specify emission regulations as a function of boiler size. Standards being studied for small and intermediate size boilers specify emission limits ranging from 1.2 to 2.0 lb $SO_2/10^6$ Btu and sulfur reduction requirements ranging from zero to 90 percent.

Prevention of Significant Deterioration

A new Part C (Sections 160-169) was incorporated into the Clean Air Act Amendments of 1977 for the prevention of significant deterioration (PSD) in the present ambient air quality. Limits on the permissible deterioration of air quality are: Class I, little or no deterioration; Class II, limited deterioration; and Class III, moderate deterioration.

Any new source in an area subject to the PSD provisions of the Act must employ the Best Available Control Technology (BACT) for each pollutant subject to regulation. BACT, which is determined on a case-by-case basis, must consider the available technologies and the energy, environmental, and economic impacts of each. Thus, the BACT identified for PSD may require higher levels of control than specified by the NSPS for that source category.

Nonattainment Areas

Provisions were also incorporated into the Clean Air Act Amendments of 1977 to alleviate air pollution problems in areas where one or more air pollutants exceed any NAAQS. In "nonattainment areas," new sources must employ pollution control technology which provides the "lowest achievable emission rate" (LAER). Before construction permits are issued, a reduction in emissions from existing sources must be obtained to more than "offset" the new source emissions. Standards for new sources in nonattainment areas are to be set by the individual states through the SIPs on a case-by-case basis.

PHYSICAL DESULFURIZATION POTENTIAL OF U.S. COALS

The decision to use a given set of control technologies as a method for complying with SO_2 emission regulations will be based on the technical applicability and relative cost of the various control options. The three control options most likely to be considered for a wide range of regulations are physical coal cleaning (PCC), flue gas desulfurization (FGD), or a combination of physical coal cleaning and flue gas desulfurization (PCC + FGD). Aside from economic factors the primary constraints which must be considered are the SO_2 emission regulation and the properties of the coal(s) which are to be used.

FGD is a flexible technology which can be used to comply with a wide range of SO_2 emission regulations. PCC, while generally less costly,is limited in its range of application because of the inherent properties of coal. An understanding of the physical desulfurization potential of U.S. coals is essential to the analysis of the use of PCC + FGD as a SO_2 emission control strategy.

The sulfur content of U.S. coals varies considerably. While 46 weight percent of the total reserve base can be identified as low-sulfur coal (coal with less than 1 percent sulfur), 21 percent ranges between 1 and 3 percent sulfur an additional 21 percent contains more than 3 percent sulfur. The sulfur content of 12 percent of the coal reserve base is unknown, largely because many coal beds have not been adequately characterized.

Sulfur appears in coal in two principal forms: organic sulfur and mineral sulfur in the form of pyrite. Organic sulfur, which comprises from 30 to 70 percent of the total sulfur content of most U.S. coals, is an integral part of the coal matrix and can only be removed by chemical modification of the coal structure.

Pyritic sulfur occurs in coal as discrete particles, often of microscopic size. Pyrite is a heavy mineral which has a specific gravity of 5.0; coal has a maximum specific gravity of only 1.7. The pyrite content of most coals can be significantly reduced by crushing and specific gravity separation. However, gravimetric separation of very fine coal and pyrite particles is not effective; separation techniques which depend on the surface or electromagnetic properties of the particles must be used.

Laboratory float-sink studies have been performed by the U.S. Bureau of Mines (USBM) on more than 455 U.S. coals to evaluate the effects of crushing and specific gravity separation on pyrite removal (Cavallaro, 1976). The samples tested were from mines in six major U.S. coal producing regions, which provide more than 70 percent of the coal used in U.S. utility boilers. In general, pyrite removal increases with decreasing coal particle size and specific gravity of separation.

The specific gravity desulfurization potential of U.S. coals varies between coal regions and between coal beds within the same region (Cavallaro, 1976). Table 1 summarizes the average sulfur values in coals from six U.S. coal regions: Northern Appalachian (NA), Southern Appalachian (SA), Alabama (A), Eastern Midwest (EMW), Western Midwest (WMW), and Western (W). Assuming that all of the pyritic sulfur could be removed by physical cleaning, average emissions from the organic sulfur would range from 0.73 to 2.86 lb $SO_2/10^6$ Btu. The percentage sulfur reduction (expressed in lb $SO_2/10^6$ Btu) achievable by removing all of the pyritic sulfur ranges from 34 to 68 percent.

The sulfur levels which could actually be achieved by crushing these coals to 3/8-inch top size and by gravimetrically separating them at 1.6 specific gravity are shown in Table 2. Total sulfur emissions would range from 0.9 to 5.5 lb $SO_2/10^6$ Btu. The percentage sulfur reduction at these cleaning conditions ranges from about 15 to 44 percent.

TABLE 1. AVERAGE SULFUR VALUES IN COALS FROM SIX U.S. COAL REGIONS[a]
($lb\ SO_2/10^6$ Btu)

REGION	Total Sulfur(S_t)	Standard Deviation[b]	Pyritic Sulfur(S_p)	Organic Sulfur(S_o)	S_p/S_t
Northern Appalachian	4.8	2.7	3.20	1.60	0.667
Southern Appalachian	1.6	1.0	0.59	1.01	0.369
Alabama	2.0	1.5	1.04	0.96	0.520
Eastern Midwest	6.5	2.1	3.80	2.70	0.585
Western Midwest	9.0	4.5	6.14	2.86	0.682
Western	1.1	0.6	0.37	0.73	0.336

[a] Cavallaro, 1976

[b] Standard deviation of total sulfur values

TABLE 2. **SUMMARY OF AVERAGE PHYSICAL DESULFURIZATION POTENTIAL OF COALS BY REGION** *

(Cumulative analysis of float 1.60 product for 3/8-inch top size)

Coal Region	No. Samples	Btu Recovery, Percent	Ash, Percent	Pyritic Sulfur, Percent	Total Sulfur, Percent	Emission on Combustion, lb $SO_2/10^6$ Btu	Calorific Content, Btu/lb
Northern Appalachian	227	92.5	8.0	0.85	1.86	2.7	13,766
Southern Appalachian	35	96.1	5.1	0.19	0.91	1.3	14,197
Alabama	10	96.4	5.8	0.49	1.16	1.7	14,264
Eastern Midwest	95	94.9	7.5	1.03	2.74	4.2	13,138
Western Midwest	44	91.7	8.3	1.80	3.59	5.5	13,209
Western	44	97.6	6.3	0.10	0.56	0.9	12,779
Total U. S.	455	93.8	7.5	0.85	2.00	3.0	13,530

* Callavaro, 1976.

The above cleaning conditions are representative of the physical desulfurization which can be obtained by applying technology now used primarily to remove mineral matter from steam coals. By optimization of physical coal cleaning processes, it is probable that from 50 to 60 percent of the total sulfur can be removed from high sulfur coals. Improvements in the cleaning conditions used for low sulfur coals could probably improve total sulfur removal capabilities to the range of 20 to 30 percent.

Estimates of the potential amount of coal in the NA, EMW, and W coal regions which could be used to comply with various emission levels are shown in Figures 1 through 3. The quantities of coal from each of these regions which can achieve an emission limit of 2.0 lb $SO_2/10^6$ Btu at varying percentages of sulfur reduction are shown in Figure 4 through 6 (Hall, 1979).

In evaluating this data and other information on U.S. coals, the following general observations can be made:

1. PCC can be used for moderate reductions in the sulfur contents of high sulfur Northern Appalachian and Midwestern coals. However, few of these coals can be cleaned to the 1.2 lb $SO_2/10^6$ Btu level specified by the current NSPS for coal fired steam generators.
2. Many Southern Appalachian, Alabama, or Western coals are capable of meeting the current NSPS coal fired steam generators, either as-mined or after cleaning.
3. Emission regulations which specify emission limits below about 1.0 lb $SO_2/10^6$ Btu preclude the use of physically cleaned high sulfur coal for compliance with these regulations. This is a consequence of the high organic sulfur contents of these coals and the fine sized pyrite which cannot be removed by PCC.
4. Emission regulations which specify sulfur reduction requirements above 30 percent preclude the use of low sulfur coals from compliance with these regulations. The percentage of sulfur which can be removed from U.S. coals by PCC is directly proportional to the ratio of pyritic to total sulfur. The fraction of pyritic

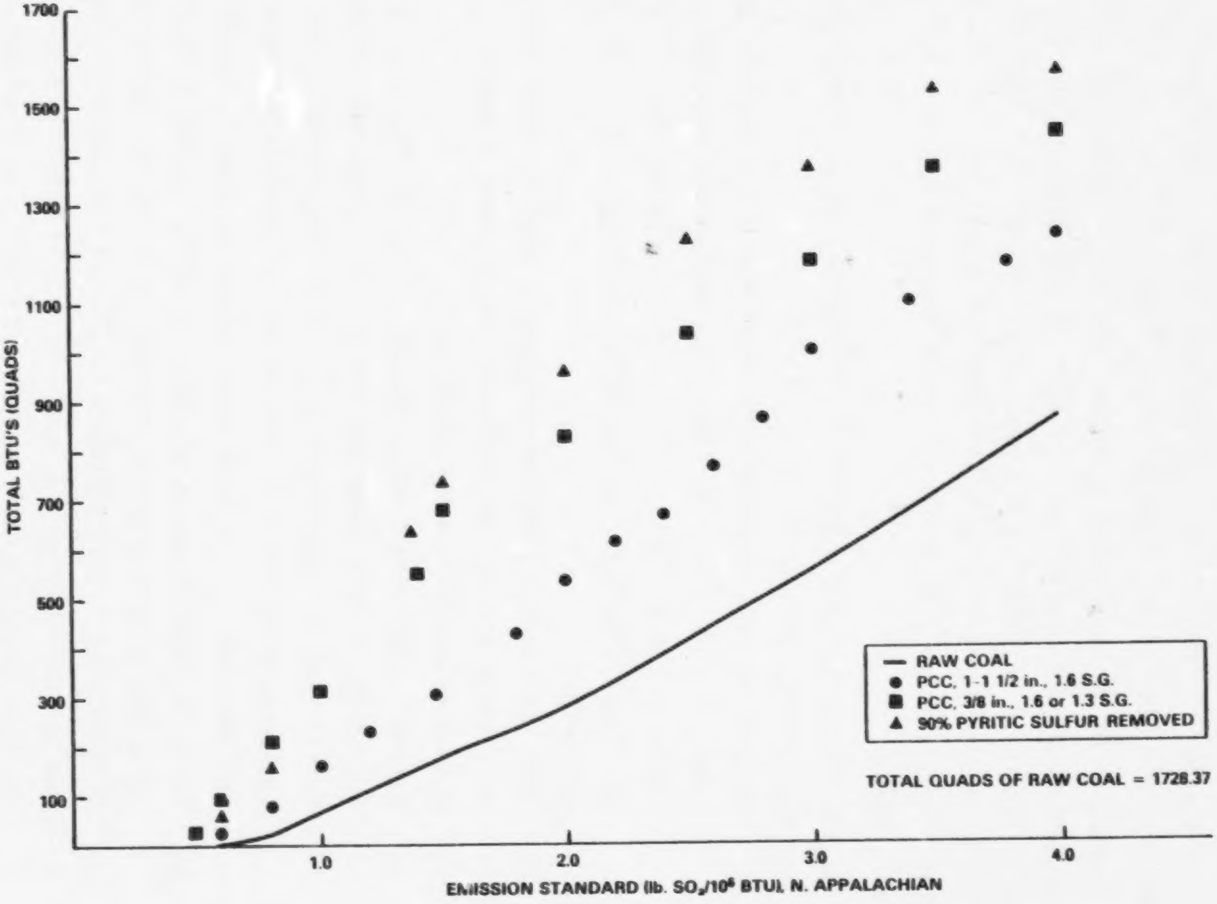

FIGURE 1 ENERGY AVAILABLE IN N. APPALACHIAN RESERVE BASE AS A FUNCTION OF EMISSION STANDARDS FOR VARIOUS PHYSICAL COAL CLEANING LEVELS

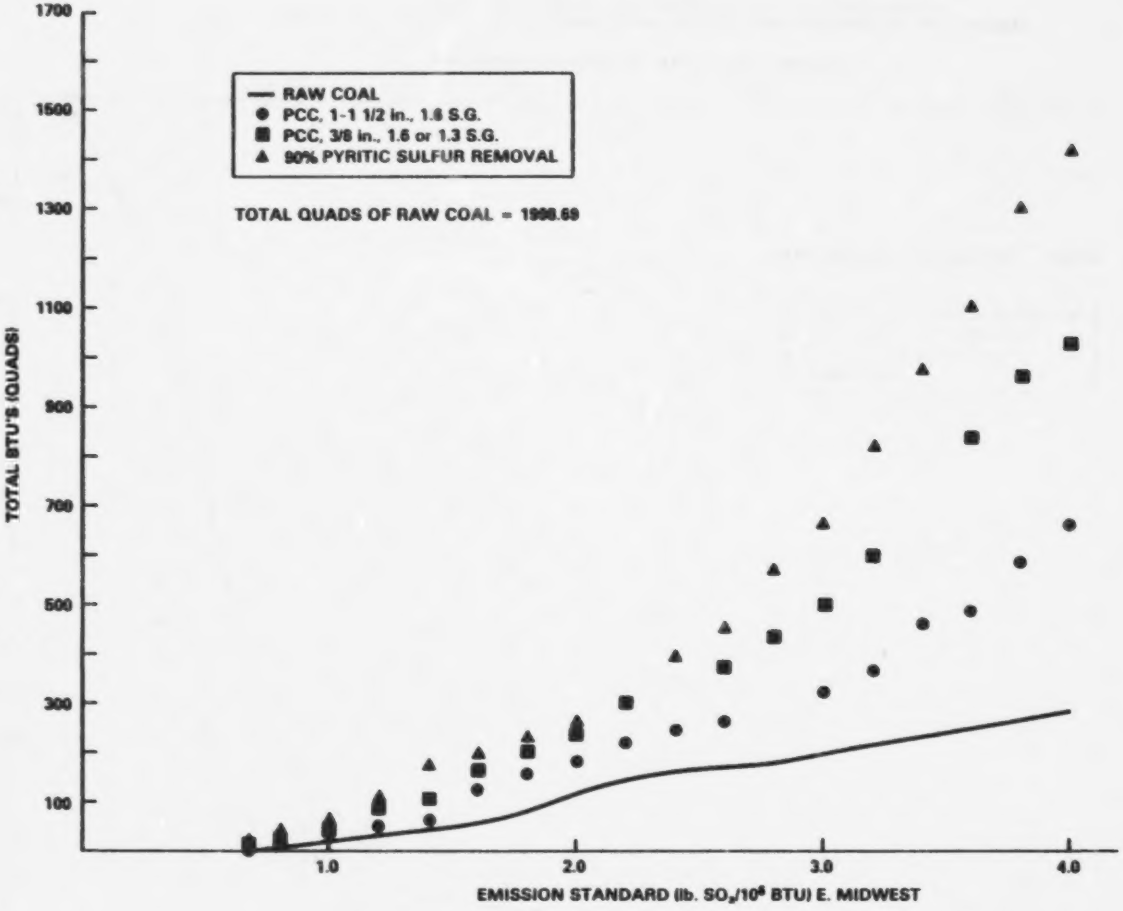

FIGURE 2 ENERGY AVAILABLE IN E. MIDWEST RESERVE BASE AS A FUNCTION OF EMISSION STANDARDS FOR VARIOUS PHYSICAL COAL CLEANING LEVELS

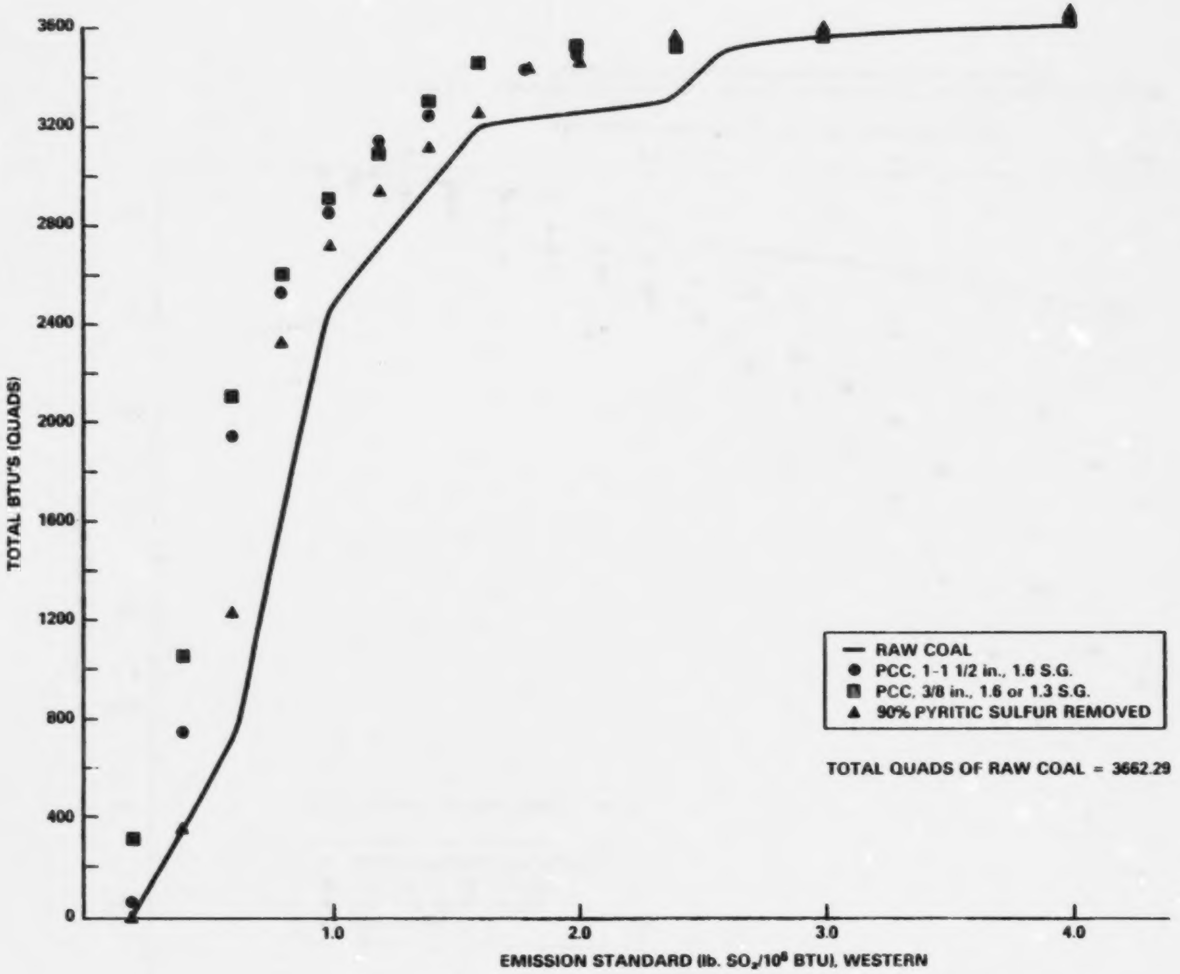

FIGURE 3 ENERGY AVAILABLE IN WESTERN RESERVE BASE AS A FUNCTION OF EMISSION STANDARDS FOR VARIOUS PHYSICAL COAL CLEANING LEVELS

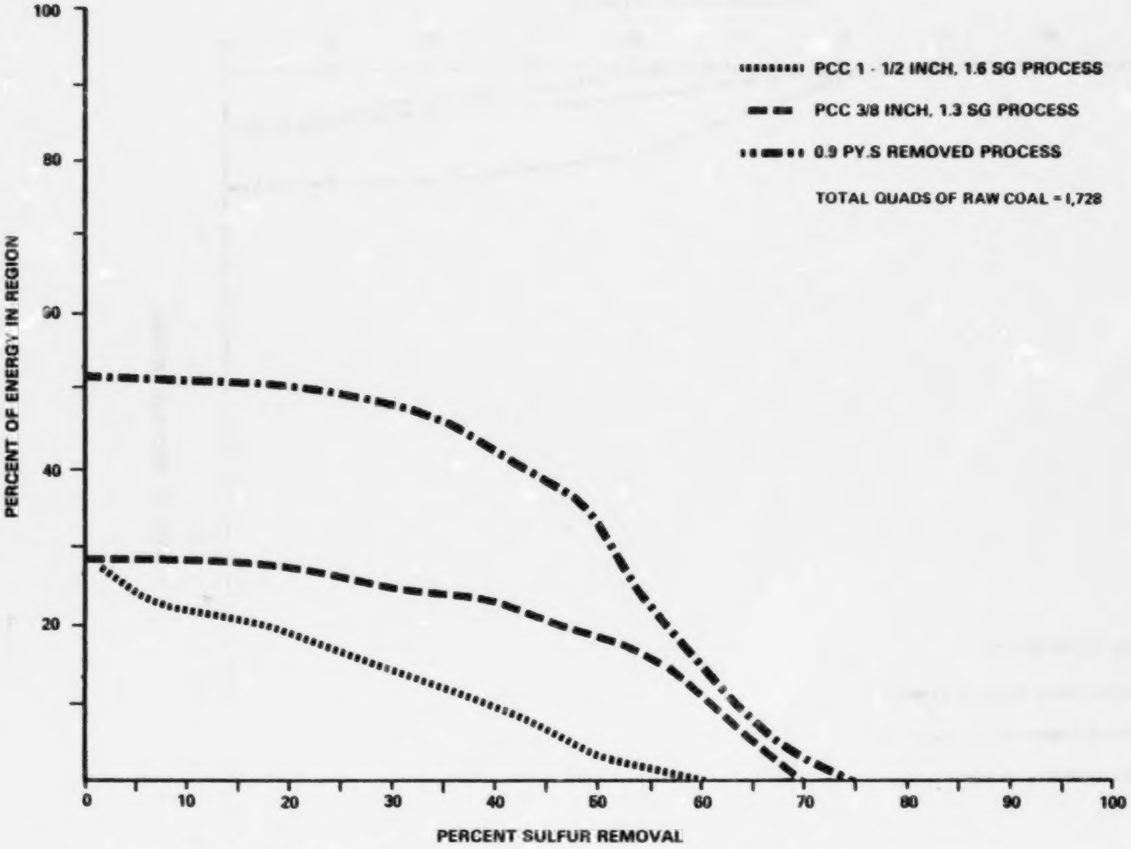

FIGURE 4. NORTHERN APPALACHIAN REGION; ENERGY AVAILABLE TO MEET PERCENT SULFUR REMOVAL STANDARDS WITH A 2.0 LB $SO_2/10^6$ BTU EMISSION LIMIT. (HALL, 1979)

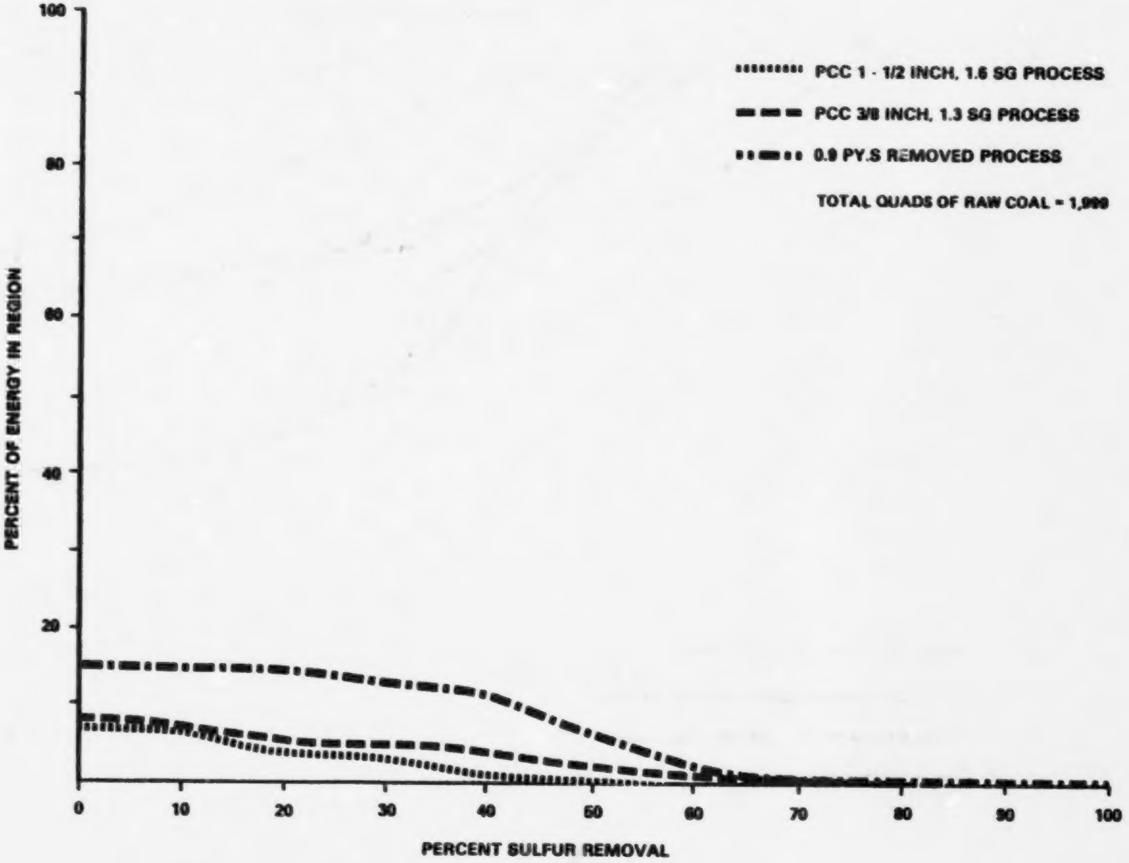

FIGURE 5 EASTERN MIDWEST REGION; ENERGY AVAILABLE TO MEET PERCENT SULFUR REMOVAL STANDARDS WITH A 2.0 LB $SO_2/10^6$ BTU EMISSION LIMIT. (HALL, 1979)

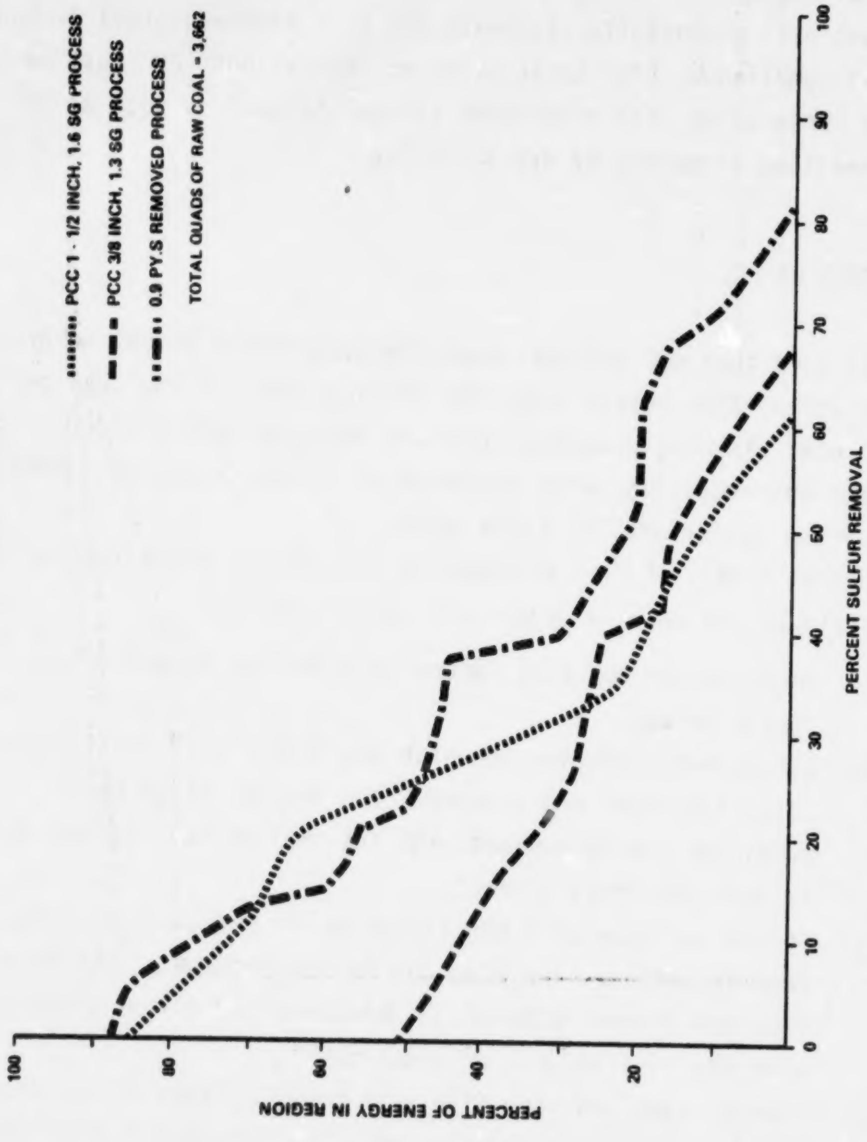

FIGURE 6. WESTERN REGION; ENERGY AVAILABILITY TO MEET PERCENT SULFUR REMOVAL STANDARDS WITH A 2.0 LB $SO_2/10^6$ BTU EMISSION LIMIT. (HALL, 1979)

sulfur in low sulfur coals is less than 40 percent. Rarely can sufficient pyrite be removed from low sulfur coals to achieve a total sulfur reduction above 30 percent.

5. Emission regulations which specify any combination of emission limit below 1.0 lb $SO_2/10^6$ Btu and sulfur reduction above 30 percent will essentially eliminate PCC as a single control technology for compliance. For these types of regulations, PCC must be used in conjunction with some other control technology such as wet limestone scrubbing or dry scrubbing.

STATUS AND COSTS OF PCC

There are more than 460 physical coal cleaning plants which can process approximately 360 million metric tons (400 million tons) of raw coal per year. The principal coal cleaning processes used are oriented toward product standardization and ash reduction, with increased attention to sulfur removal as the demand for low sulfur utility fuels grows.

Five general levels of coal preparation are used in upgrading coal. Each level includes one or more major unit operations:

Level 1 Breaking for top size control and for the removal of coarse refuse.

Level 2 Coarse beneficiation in which the larger coal particles (plus 3/8-inch) are treated. The treated large coal particles are recombined with the smaller coal particles to form the final product.

Level 3 Coarse and fine size beneficiation in which all of the feed is wetted. The plus 28M is beneficiated. The very fine 28M x zero material is dewatered and either shipped with the cleaned coal or discarded.

Level 4 Coarse, fine and very fine size beneficiation in which all of the feed is wetted and cleaned. The 1/4-inch x zero fraction is generally dried to limit the moisture content.

Level 5 Full beneficiation for optimal ash and sulfur rejection. This may involve crushing the coal to finer sizes and producing a number of coal products, each with a different ash and sulfur content.

Coal cleaning costs are sensitive to plant capacity, plant complexity (level of cleaning) and coal replacement costs. Coal replacement costs are defined as the cost of coal energy which must be discarded with the plant residue.

The percentage of coal energy (Btu's) recovered at modern coal preparation plants is generally greater than 90 percent. With current high coal prices, the annualized costs of physical coal preparation are more sensitive to coal replacement costs than to plant complexity costs (Kilgroe, 1977). Removal of finely distributed pyrite from coal entails a high degree of complexity. A high degree of complexity is also required to recover a large fraction of the fine coal which at times has been discarded by steam coal plants. Thus increased plant complexity for pyrite removal is consistent with trends to minimize annual costs by increasing the recovery of fine coal.

Total coal preparation costs exclusive of coal replacement costs may range from \$0.04 to \$0.18/10^6 Btu (Buder, 1977, and Holt, 1978). For an 11,000 Btu/lb raw coal costing \$20/ton, coal replacement costs from a plant with 95 percent Btu recovery would be \$0.046/10^6 Btu. Typical costs for coal preparation are shown in Table 3 (Buder, 1977).

STATUS AND COSTS OF FGD

FGD systems are capable of removing more than 90 percent of the SO_2 from the flue gas combustion products of coals with a wide range of sulfur content. Annualized control costs increase with the increased amounts of sulfur removed. In a given size boiler the costs of removing a fixed percentage of sulfur from the flue gas stream are greater for high sulfur coals than for low sulfur coals. Although total control costs increase with the volume of flue gas cleaned, normalized control costs for small to moderate sized boilers are greater than for large boilers. Figure 7 illustrates the effects of boiler size and sul-

Table 3. Capital and Operating Cost Summary For Various Coal Preparation Plants*

Cleaning Level	1 [f]	2	3	4	5
Coal	Montana (Rosebud)	W. Virginia (Cedar Grove)	Colorado (Montose City)	Pennsylvania (Lower Kittanning)	Pennsylvania (Upper Freeport)
Plant Construction Cost	6,800,000	12,100,000	20,300,000	24,100,000	37,800,000
Pre-construction and Owners Costs	1,907,000	2,651,000	5,485,559	6,562,000	10,250,500
Total Depreciable Capital Cost	8,707,000	14,751,000	25,785,559	30,662,000	48,050,500
Land Cost	150,000	225,000	330,000	900,000	600,000
Total Capital Costs	8,857,000	14,976,000	26,115,559	31,562,000	48,650,500
Annual Costs[a,b]					
Operating and Maintenance	904,851	1,303,067	1,799,678	2,579,111	3,896,406
Capital-Related[c]	993,767	1,674,523	2,735,864	3,380,276	5,285,128
Total Annual Costs	1,898,618	2,977,596	4,535,542	5,959,387	9,181,534
Return on Investment[d]	797,130	1,327,590	2,359,400	2,840,580	4,378,545
Coal Processing Costs[e]					
$/ton dry product	0.584	0.916	1.396	1.834	2.825
$/10⁶ Btu dry product	0.026	0.036	0.053	0.064	0.097
Return on Investment					
$/ton dry product	0.245	0.408	0.723	0.874	1.347
$/10⁶ Btu dry product	0.011	0.016	0.028	0.031	0.046
Total Costs					
$/ton dry product	0.829	1.324	2.119	2.708	4.172
$/10⁶ Btu dry product	0.037	0.052	0.081	0.095	0.143

* (Buder, 1977)

[a] Corresponds to 3.25 million tons per year (dry) at 250 thirteen-hour annual operating days.

[b] Excludes coal replacement costs.

[c] Calculated at a 7:3 debt/equity ratio repaying debt with 9% 20-year bonds.

[d] Calculated on equity with a before tax return of 30% with no discounting.

[e] All costs for Case 5 reflect the combined middlings and clean coal products. The separation of these costs will depend on market conditions.

[f] Excluding thermal drying option which would add $0.45 per ton dry product to the processing cost, and $6,900,000 to the capital investment.

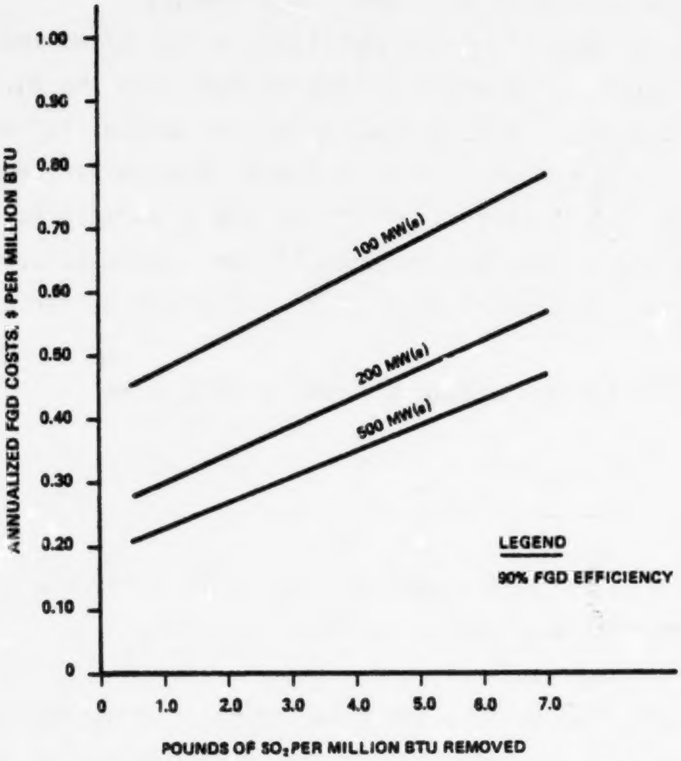

FIGURE 7 ESTIMATED COST OF LIME/LIMESTONE FGD SYSTEM (McGLAMERY, 1975 AND ISAACS, 1977)

fur removal requirements on costs for a lime/limestone FGD system. These costs were developed with the use of FGD cost data generated from computerized cost models (McGlamery, 1975,and Isaacs, 1977).

Partial scrubbing can be used to reduce control costs in cases where full scrubbing is not required. In partial scrubbing, part of the flue gas stream bypasses the scrubber. This reduces the amount of flue gas to be treated and reduces or eliminates flue gas reheat requirements.

Partial scrubbing can be used if an emission limit is specified and the fuel sulfur value is low enough to permit treatment of less than the full flue gas stream. Alternatively, if a high level of sulfur control is required and precombustion sulfur removal credit is given, then sulfur removal by coal cleaning techniques can be used to permit the use of partial scrubbing. Partial scrubbing, of course, is only advantageous if the reduced scrubber costs more than offset any higher costs attributed to the use of low sulfur coals or coal cleaning.

Only lime/limestone FGD systems are considered in this paper.

COST TRADE-OFFS

With combinations of PCC + FGD, a number of cost trade-offs must be considered in assessing the cost impacts of clean coal use. The use of cleaned coal for power generation may impact: the design and operation of the boiler, the flue gas treatment equipment,or other components of the overall energy generation system. Sulfur variability,which can have a substantial impact on FGD costs,is discussed separately.

Boiler Cost Impacts

Some of the most important considerations in the design of a boiler are the characteristics of a coal and its ash. Reliable boiler operation depends on the application of design techniques utilized to minimize slagging, fouling, and corrosion problems. These problems in a large measure directly affect boiler availability. Of the reasons advanced for the use of cleaned coal, perhaps the greatest single benefit to be obtained (other than the control of emissions) is in the area of fireside performance. Fireside problems are responsible for many coal fired operational difficulties resulting in both forced and scheduled outages. They significantly affect the cost of boiler operation and maintenance, the capacity factor, and (in the case of utilities) the availability of the generating facility. By modifying the coal characteristics which contribute to these problems, coal cleaning can favorably affect the economical use of coal. Coal characteristics vary widely, however, and different methods of cleaning can have diverse effects on the properties of coal ash and sulfur content. For this reason, the effects of coal cleaning on boiler operation will, in a given circumstance, depend on both the cleaning method used and the original characteristics of the coal.

The net effect of coal cleaning on the operating and maintenance costs of boilers (as related to slagging, fouling, and corrosion) is difficult to quantify because of the many variables involved. In a previous study on coal cleaning and scrubbing (Hoffman, 1976), boiler maintenance cost savings were postulated to be related to the additive reduction in the sulfur and ash content of the coal (see Table 4). These cost savings were based on TVA studies on the effects of coal quality on the operation and maintenance of large central station boilers (Holmes, 1969).

TABLE 4.*
THE EFFECT OF ASH AND SULFUR CONTENT ON BOILER MAINTENANCE COSTS

Additive Reduction in Ash and Sulfur, percent	Maintenance Cost Savings, $/ton Coal
>15	0.33
12 - 15	0.30
9 - 12	0.27
7 - 9	0.24
5 - 7	0.20
3 - 5	0.17
2 - 3	0.13

* (Hoffman, 1976)

Recent papers by Cole and Phillips postulate a number of cost penalties which may be attributed to the total ash and sulfur content of coal (Cole, 1978, and Phillips, 1979). These penalties include increased maintenance costs, loss of peaking capacity and reduction in plant availability. Total cost penalties at 15 and 17.5 percent ash plus sulfur are given as $0.38/ton and $0.75 ton (see Table 5). Above 17.5 percent ash and sulfur the postulated total cost penalties rise exponentially, reaching a value of $6.41/ton at 25 percent ash and sulfur.

TABLE 5.*

COST PENALTIES ASSOCIATED WITH ASH AND SULFUR CONTENT OF THE COAL

Ash Content, percent	Sulfur Content, percent	Total A&S percent	Cost Penalty, $/ton Coal Fired Maintenance Costs	Peaking Capacity	Rated Capacity	Plant Availability	Total
10.5	2.0	12.5	0	0	0	0	0
12.5	2.5	15.0	0.38	0	0	0	0.38
14.5	3.0	17.5	0.75	0	0	0	0.75
16.5	3.5	20.0	1.13	0.19	1.08	0.47	2.87
18.5	4.0	22.5	1.50	0.23	2.08	0.91	4.72
20.5	4.5	25.0	1.88	0.21	3.00	1.32	6.41

* (Phillips, 1979)

The amount of data which can be used to correlate the effects of PCC and boiler operating and maintenance costs is extremely limited. Studies are now underway by EPA and EPRI to develop improved correlations between coal quality and operating costs. Until additional information becomes available the studies cited above can be used to estimate the range of boiler related cost benefits which may be available with improved coal quality.

FGD Cost Impacts

If the full flue gas stream is scrubbed then it is necessary to reheat it to avoid severe stack corrosion problems and ensure proper dispersion of the stack plume into the atmosphere. The use of partial scrubbing can reduce or eliminate reheater capital and operating costs. In addition, after reheat requirements have been met, the use of further bypass will lower costs by reducing the volume of flue gases scrubbed. A study of energy requirements of a limestone FGD system concluded that for "any fixed set of coal and plant characteristics, FGD energy use is minimized by operating the scrubber at high efficiency (90-93%) while bypassing as much flue gas as permitted by the applicable emission standard". An economic analysis of the base case plant design further indicated that FGD capital costs as well as operating costs were reduced by partial bypass (Rubin, 1978).

The amount of partial scrubbing which can be made available by coal cleaning depends upon the sulfur removal efficiencies of the coal cleaning and scrubbing processes and upon the applicable emission regulations which must be met. Figure 8 presents the amounts of flue gas which may be bypassed for a number of coal cleaning sulfur reduction efficiencies and total sulfur emission reduction requirements (a 90 percent FGD sulfur removal efficiency is assumed). For a 90 percent sulfur removal standard the available bypass would probably not be sufficient to meet total reheat requirements, even at high coal cleaning sulfur efficiencies. If the scrubber efficiencies are raised to 95 percent and the coal cleaning sulfur-removed efficiencies are 30 percent or larger, then the amount of bypass available may be sufficient for reheat requirements (see Figure 9).

Emission regulations which contain an emission limit below which additional control is not required can use naturally occurring low sulfur coals or cleaned coals to permit a high amount of bypass. For example: cleaning a 2.0 lb $SO/_2/10^6$ Btu coal to remove 20 percent of the coal sulfur would

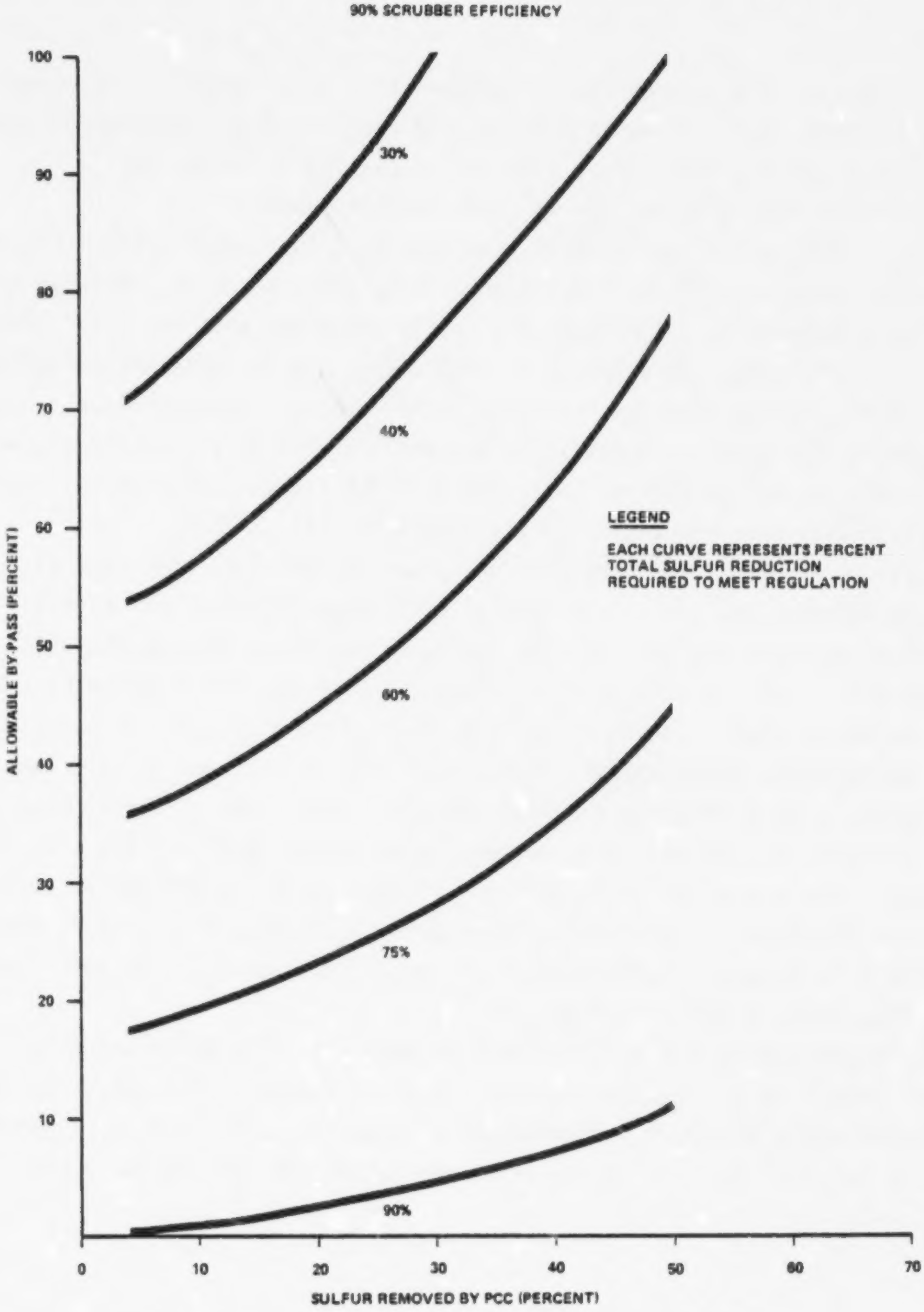

FIGURE 8 EFFECT OF PCC SULFUR REMOVAL EFFICIENCY ON ALLOWABLE FGD BY-PASS FOR 90% SCRUBBER EFFICIENCY

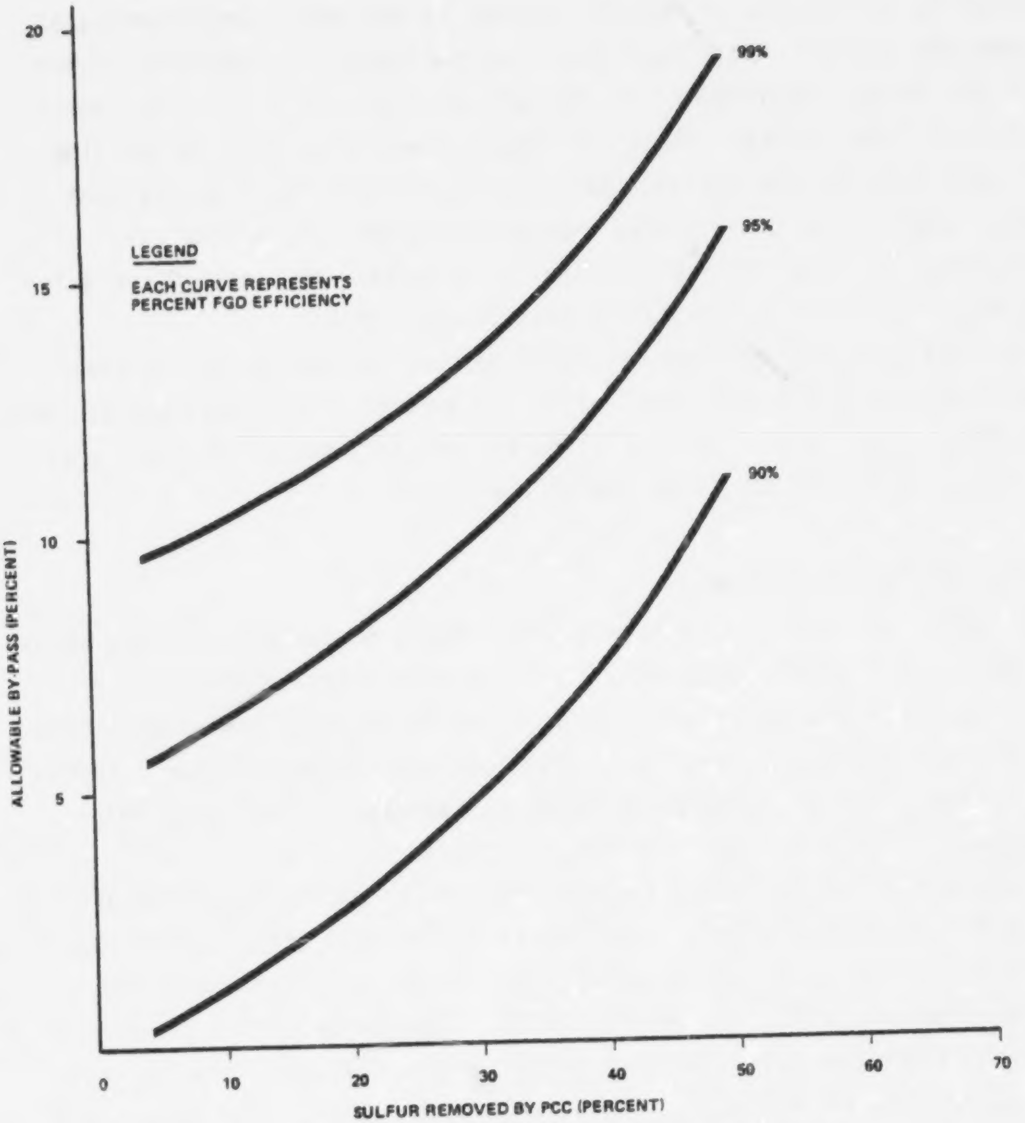

FIGURE 9 EFFECT OF PCC AND FGD REMOVAL EFFICIENCIES ON ALLOWABLE FGD BY-PASS FOR 90% SULFUR REDUCTION STANDARD

allow for a 23.6 percent bypass where a 90 percent efficiency scrubber was used to meet a 0.5 lb $SO_2/10^6$ Btu emission limit (75 percent total sulfur control).

The potential energy cost saving available by an allowable bypass may be estimated by the results of Rubin's studies on the energy requirements of a limestone FGD system. Rubin found that for the range of parameters tested, the total FGD energy requirement was equivalent to between 2.5 and 6.1 percent of the total power plant energy output (or input) when 100 percent of the flue gas was treated in the FGD system. Sensitivity analyses for a 3.5 percent sulfur coal used in the study showed that treating the entire flue gas stream required 10 to 30 percent more energy to achieve the same SO_2 emission standard than a system with partial by pass (Rubin, 1978).

As a first approximation one can conclude that the energy saving available from bypass could range from 0.25 to 1.8 percent of the coal energy input to the boiler. For a coal costing $\$1.00/10^6$ Btu this would represent a cost savings of 2.5 to 18 mills/million Btu of heat input.

Sulfur Variability Cost Impacts

Fuel sulfur variability can have a large impact on the costs of SO_2 emission control. Many current regulations specify that sulfur emissions are "never" to exceed a stated emission limit. Compliance with these regulations is generally determined by one or more emission tests conducted over a specific time period. This time period provides an "averaging time" over which average emissions cannot exceed the emission limit.

The amount of coal burned in a given size boiler with a specified averaging time is the characteristic coal "lot size" for that boiler. Previous studies indicate that coal sulfur variability probably increases with decreasing lot size (Nelson, 1977, and Versar, 1979). Compliance with a given regulation is therefore more difficult for small boilers than for large boilers.

The sulfur content of coal varies between coal regions, coal seams, and locally within each coal mine. The extreme nature of coal sulfur variability is illustrated by Figure 10, which shows the sulfur content of Helen Mine coal for successive mining days over a 5 year period. As suggested by Figure 9,sulfur variation over short and long time periods is of concern.

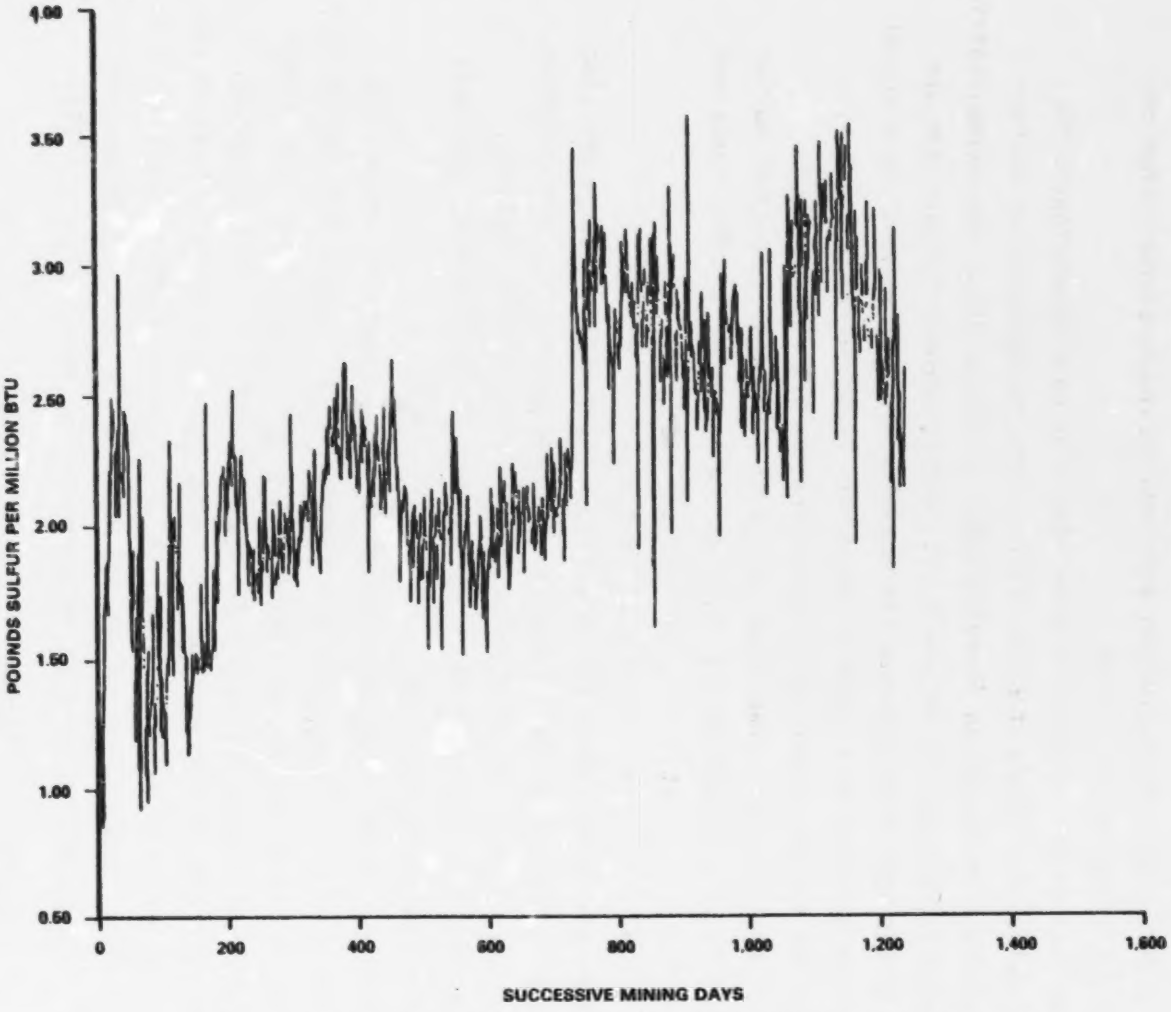

FIGURE 10 POUNDS OF SULFUR PER MILLION BTU FOR SUCCESSIVE MINING DAYS FOR HELEN MINE, JANUARY 1970 THROUGH DECEMBER 1975. (THOMAS, 1978)

In designing a sulfur emission control system,provisions must be made for compliance with the emission averaging time and the maximum mean sulfur content of the coal which is to be used.

Data on coal sulfur variability show distributions skewed toward the the higher coal sulfur levels (Nelson, 1977). This skewness can be characterized by either log-normal or inverted gamma distributions. The skewed distribution is especially apparent for small lot sizes: bore hole samples are the smallest. As the lot sizes increase the skewness is decreased and the distribution can be approximated by a gaussian or normal distributions.*

Coal sulfur values exhibit different variances over hourly (σ_h^2), weekly (σ_w^2), monthly (σ_m^2), and yearly (σ_y^2) time periods. The total sulfur variability will be composed of the short and long term variance as described by the equation:

$$\sigma^2 = \sigma_h^2 + \sigma_d^2 + \sigma_w^2 + \sigma_m^2 + \sigma_y^2$$

The time related variances for coal from a given mining reserve are also related to the manner in which the coal is mined or used; i.e., the variation is spatially correlated. If mining is conducted along sulfur isolines, daily and weekly variances will be small. If coal is mined across isolines, short term variances may be large.

Emission regulations with long averaging times (30 days or longer) can mitigate the effects of short term sulfur variability. Longer term variability, which can be characterized by the change in the monthly or yearly mean sulfur values, must be accounted for in the design of the emission control system.

It is convenient to characterize the variability of coal sulfur values and stack emission values in terms of a mean value (μ) and a standard deviation (σ). The extremes in the coal sulfur values and stack sulfur emission values (see Figure 10) can then be described by equations of the form (Kilgroe, 1979):

$$\text{Coal sulfur extremes} = \mu_c\ (1 \pm a_c RSD_c)$$

$$\text{Stack sulfur emission extremes} = \mu_s\ (1 \pm a_s RSD_s)$$

where RSD = σ/μ is the coefficient of variation or the relative standard deviation. The product,a RSD,is defined as the coal variability factor; $a_c RSD_c$ is the coal variability factor and a_sRSD_s is the scrubber variability factor. The coefficient "a" is used as needed to express a confidence interval relating to a given lot size or averaging time.

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* Classical statistics do not adequately describe the variance properties of coal. Geostatistics, a recent development which is capable of describing distributions with a spatial correlation of properties,appears to offer a better technique for characterizing coal sulfur variability (Thomas, 1978).

Design constraints placed on the FGD by sulfur variability may be evaluated in terms of the mean sulfur reduction requirement of the emission standard ($\bar{\eta}$). The percentage removal requirement must hold for all cases, the extremes as well as the mean. The required removal conditions at the mean are:

$$\bar{\eta} = 1 - \frac{\mu_s}{\mu_c}$$

The removal efficiency at the extremes provides bounds for all other cases. The four extreme cases illustrated in Figure11 can be represented by the equation:

$$\bar{\eta} = 1 - \frac{\mu_s}{\mu_c} \frac{(1 \pm a_sRSD_s)}{(1 \pm a_cRSD_c)} = 1 - (1-\bar{\eta}) \frac{(1 \pm a_sRSD_s)}{(1 \pm a_cRSD_c)}$$

Table 6 presents data showing the required efficiency at extreme conditions corresponding to selected coal and sulfur variability factors. All cases correspond to a mean FGD sulfur removal requirement of 85 percent. Case III is not presented because the required efficiency at these extremes is less than those at mean conditions.

In the worst case, Case IV, the effects of coal and scrubber variability are compounded. At $a_cRSD_c = 0.10$ and $a_sRSD_s = 0.8$ the required design removal efficiency at the extreme is 97.3 percent. For values of coal sulfur variability which may be expected from uncleaned or unmixed coals ($a_cRSD_c = 0.5$) and for the upper range of scrubber sulfur variability ($a_sRSD_s = 0.8$) the required design removal efficiency at the extreme is 98.0 percent.

Mixing and preparation may be used to reduce variations in coal properties. Mixing of coal to form a more homogeneous product will occur during mining, handling, blending,and preparation. During mining, coal is often removed from multiple mining faces during the same time period. Coal from these locations is combined to form a single run-of-mine (ROM) product. Coal from each of the faces may have a different mean sulfur value and

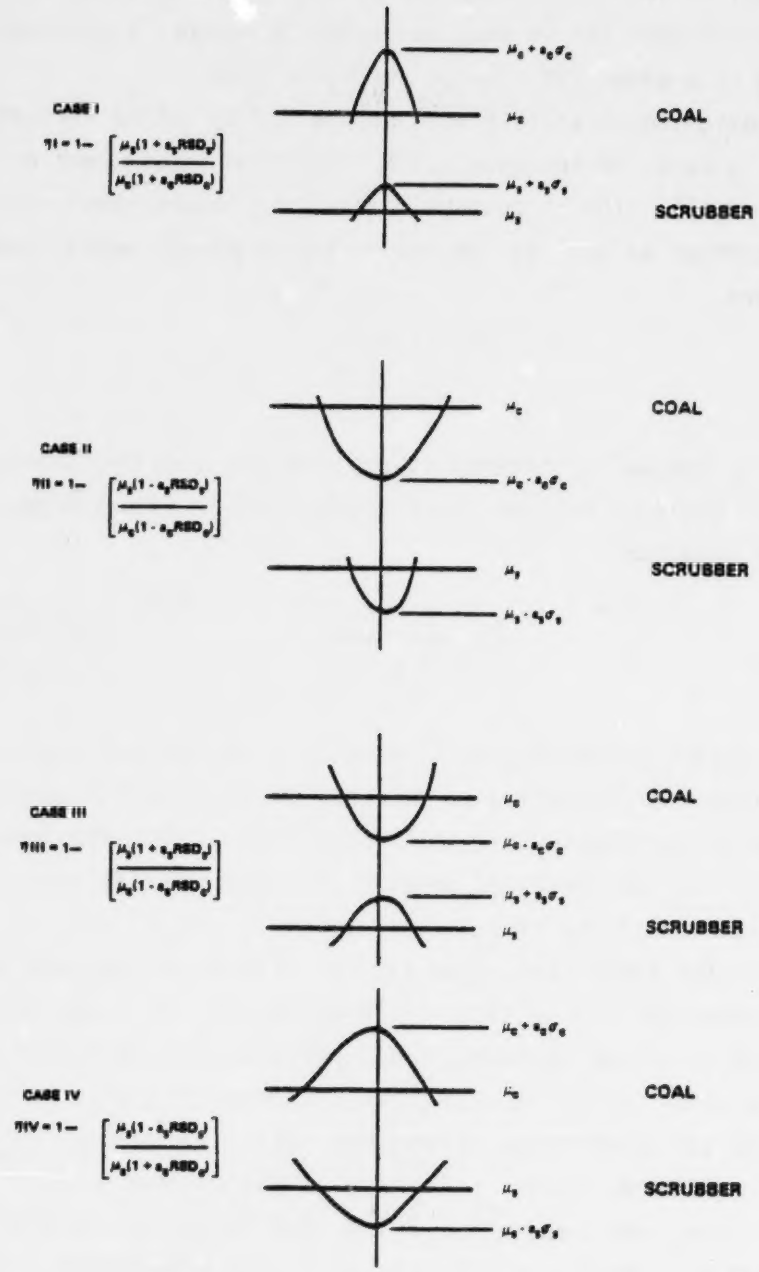
CASE I
$\eta I = 1 - \left[\frac{\mu_s(1 + a_s RSD_s)}{\mu_c(1 + a_c RSD_c)}\right]$
$\mu_c + a_c\sigma_c$
μ_c
COAL
$\mu_s + a_s\sigma_s$
μ_s
SCRUBBER
CASE II
$\eta II = 1 - \left[\frac{\mu_s(1 - a_s RSD_s)}{\mu_c(1 - a_c RSD_c)}\right]$
μ_c
COAL
$\mu_c - a_c\sigma_c$
μ_s
SCRUBBER
$\mu_s - a_s\sigma_s$
CASE III
$\eta III = 1 - \left[\frac{\mu_s(1 + a_s RSD_s)}{\mu_c(1 - a_c RSD_c)}\right]$
μ_c
COAL
$\mu_c - a_c\sigma_c$
$\mu_s + a_s\sigma_s$
μ_s
SCRUBBER
CASE IV
$\eta IV = 1 - \left[\frac{\mu_s(1 - a_s RSD_s)}{\mu_c(1 + a_c RSD_c)}\right]$
$\mu_c + a_c\sigma_c$
μ_c
COAL
μ_s
SCRUBBER
$\mu_s - a_s\sigma_s$

FIGURE 11. COAL AND FGD SULFUR VARIABILITY CASES

TABLE 6. EFFECT OF SULFUR VARIABILITY FACTORS UPON VARIABLE EFFICIENCY[a]

Extreme[b]	Coal Variability Factor (a_cRSD_c)	FGD Variability Factor (a_sRSD_s)	Required Efficiency at Extreme (percent)
Case I	0.30	0.00	88.5
		0.10	87.3
		0.20	86.2
	0.50	0.10	89.0
		0.40	86.0
Case II	0.10	0.20	86.7
		0.40	90.0
		0.80	96.7
	0.30	0.40	87.1
Case IV	0.10	0.20	89.1
		0.40	91.8
		0.80	97.3
	0.30	0.10	89.6
		0.20	90.8
		0.40	93.1
		0.80	97.7
	0.50	0.10	91.0
		0.80	98.0

(a) For mean removal efficiency of 85 percent

(b) Case III not presented because required efficiency at these extremes less than those at mean conditions.

a different degree of sulfur variability. The mining methods and the way in which coal is combined from the different faces will have a significant effect on the mean sulfur values and the sulfur variability.

The purposeful mixing of coal from two or more mines to reduce variability or modify average coal properties can also be accomplished by "blending."

Mixing during handling, storage,and transportation will likely result in a more homogeneous product with reduced variability.

Data which provide quantitative information correlating the reduction in sulfur variability to the individual mining, handling and blending operations are not currently available. However data sets from coal suppliers and utility sources indicate that RSDs for large lot sizes of ROM coal may range from approximately 10 to 30 percent (Nelson, 1977). Data from coal preparation plants suggest that the product coal RSDs may normally be in the range of 5 to 10 percent (Versar, 1979).

Limited data are available on the variability of sulfur emissions from FGD scrubbers on utility boilers (Kelly, 1978). The RSD values for three FGD units evaluated by OAQPS for 3 and 24 hour averaging times normally fell in the range of 20 to 45 percent. A fourth highly efficient FGD unit (approximately 95 percent removal efficiency) exhibited outlet RSD values from 70 to 104. These high RSD values resulted primarily from extremely low mean emission values which ranged from 0.233 to 0.273 lb $SO_2/10^6$ Btu. These data suggest a probable range of FGD sulfur variability factors ranging from 40 to 100 (value of $a_s = 2.0$ is assumed). Improved FGD controls should substantially reduce this variability.

The above data suggest that FGD capital and operating costs will depend on the coal sulfur variability, the emission regulation averaging time,and the ability of the FGD system to control sulfur outlet concentration , whether caused by coal sulfur variances or the basic variance of FGD process variables. Given the constraints of the regulation averaging time, variances in the coal sulfur will probably cause a larger cost impact than those cost factors related to FGD process variables (Kilgroe, 1979). Many FGD cost components are related to the total sulfur which must be removed from the flue gas stream ($\mu_c - \mu_s$). As the peak sulfur levels increase,the FGD sulfur removal capacity must be increased to account for the higher sulfur removal requirements. This increase,expressed as the ratio of costs with variability to the costs without variability,may be approximated by the expression:

$$\frac{C}{\bar{C}_s} = \frac{1}{\bar{\eta}} (1 + a_c RSD_c) \; - \frac{(1-\bar{\eta})}{\bar{\eta}} (1 - a_s RSD_s)$$

Solution of this equation for a mean sulfur removal requirement of 90 percent yields the fact that a 20 percent increase in the maximum peak coal sulfur levels will result in an increase of approximately 20 percent in the FGD costs related to amount of sulfur removed. An 80 percent increase in peak coal sulfur values would raise the sulfur related FGD costs approximately 80 percent (Kilgroe, 1979).

The cost penalties for FGD designs which incorporate a sufficient safety margin to account for both short and long term sulfur variabilities may be quite high. The use of coal blending or coal preparation may provide cost effective alternatives. Additional studies are obviously needed to evaluate these trade-offs.

ESP Cost Impacts

The removal of fly ash from flue gas is typically done by 99.5 percent efficient ESP units, usually placed downstream of the air preheaters near the boilers. The ash and sulfur contents of coals are major influences on ESP capital and operating costs. The ESP size and collection efficiency are affected by the coal sulfur content and inlet ash loading. Increases in ESP size requirements which may result from coal desulfurization may be offset by decreased ash disposal costs or increased efficiencies resulting from reduced fly ash loading. The costs of controlling particulate emissions from cleaned and uncleaned coals may not be substantially different for most cases. Studies to identify the sensitivity of particle collection to coal cleaning effects on a number of representative coals are needed to confirm this postulation.

System Cost Impacts

The primary system benefits which may accrue from the use of cleaned coal include reduced transportation costs, reduced pulverizer costs, reduced boiler ash disposal costs,and reduced mine labor costs.

Transportation and mine labor cost reductions occur if the coal is cleaned at the mine site. Cleaning reduces the ash content of the coal and increases its calorific value. This reduces the weight of coal needed to meet boiler energy requirements. A reduction in coal shipped from the mine site also reduces the amount paid to the UMW Pension and Benefit Trust Fund ($1.38/ton coal shipped to the consumer).

Pulverizer operating costs will generally be reduced in proportion to reduction in coal weight burned. While removal of the coal ash by cleaning may provide benefits beyond those attributable to the reduction coal weight pulverized, these extra benefits cannot be generally specified.

Reductions in transportation, ash disposal, pulverizer, and union trust fund costs, as estimated by Hoffman, are presented in Table 7 (Hoffman, 1976). The total average cost savings for the 12 mine and power plant site pairs evaluated were $0.56/ton. Transportation and ash disposal savings were approximately 4 to 10 times higher than pulverizer and union trust fund savings. The maximum total savings of $1.09/ton (approximately $0.04/10^6 Btu) would provide a substantial credit against the costs of coal cleaning.

TABLE 7. SYSTEM COST BENEFITS

	System Cost Benefits ($/ton) (a)		
	Minimum	Maximum	Average
Transportation Savings (b)	0.15	0.52	0.29
Ash Disposal Savings	0.10	0.36	0.19
Pulverizer Savings	0.01	0.04	0.03
UMW Trust Fund Savings (c)	0.02	0.17	0.05
Total	0.28	1.09	0.56

(a) Hoffman, 1976

(b) Average transportation distance for 12 mine power plant site pairs was 436 miles.

(c) Based on costs of $0.74

COST COMPARISONS

Economic comparisons of alternative pollution control options are complex. Factors unique to a given site often determine which option is the most cost effective.

Several comprehensive studies under the sponsorship of EPA, DOE, and EPRI are now in progress to evaluate the relative costs of FGD + PCC and FGD as options for complying with SO_2 emission regulations. The results of these studies are not expected to be available for several months. However, one may in some part anticipate the results of the studies by using information available in the literature to identify the instances where combinations of coal cleaning and FGD are likely to be economically competitive. It is emphasized that the results are only semi-quantitative in that they indicate the relative importance of major control system design factors which influence the costs of compliance with a given regulation. To assess the actual control costs and determine the most cost effective means of compliance require detailed studies which account for site specific factors which influence the costs of compliance.

Analysis Method

Cost comparisons for controlling SO_2 emissions were made using existing data on the costs of FGD and coal cleaning (see Table 8). These comparisons were limited to the control of SO_2 emissions from 500 MW boilers. The comparisons considered four different coal sulfur levels and two or three emission regulations. The coals were assumed to have compositions and properties similar to "average" coals from the Northern Appalachian, Alabama, Eastern Midwestern and Western coal regions as specified in the U.S. Bureau of Mines Publication on the Sulfur Reduction Potential of U.S. Coals (Cavallaro, 1976). Sulfur emission regulations considered included those requiring emission limits of 2.5, 2.0, 1.2, 0.8 and 0.5 lb $SO_2/10^6$ Btu or (alternatively) a 90 percent sulfur reduction.

FGD costs were determined from Figure 7 by use of the FGD sulfur removal requirement and the equivalent scrubber capacity in MW. All FGD units were assumed to operate at 90 percent efficiency. For the PCC + FGD cases the equivalent scrubbing capacity was determined by calculation of the allowable bypass using the clean coal sulfur level and required sulfur emission limit.

Table 8. PCC + FGD and PCC COST COMPARISONS

	Emission Control Level			Emission Control Level		
	1	2	3	1	2	3
I. A. Coal Region		Northern Appalachian			Alabama	
B. Raw Coal Sulfur Level, lb $SO_2/10^6$ Btu	4.8	4.8	4.8	2.0	2.0	2.0
C. Allowable Emission Limit, lb $SO_2/10^6$ Btu	2.0	1.2	0.48	1.2	0.8	0.20
D. Sulfur Removal Required, %	47.9	75.0	90.0	40.0	60.0	90.0
II. FGD Control Costs						
A. FGD Sulfur Removal Requirement, lb $SO_2/10^6$ Btu	2.3	3.6	4.32	0.4	0.8	1.44
B. Cost of 500 MW FGD at 90% Removal Efficiency, $\$/10^6$ Btu	0.28	0.34	0.37	0.21	0.220.27	
III. PCC + FGD Control Costs						
A. Sulfur Removal by PCC[a], lb $SO_2/10^6$ Btu	2.1	2.1	2.1	0.3	0.3	0.3
B. Sulfur To Be Removed by FGD, lb $SO_2/10^6$ Btu	0.7	1.5	2.22	0.5	0.9	1.5
C. FGD Sulfur Removal Requirement, %	25.9	55.6	82.2	29.4	52.9	88.2
D. Equivalent FGD Size at 90% Efficiency, MW	144	309	457	163	294	490
E. FGD Costs at Reduced Size and SO_2 Removal Requirements, $\$/10^6$ Btu	0.09	0.18	0.26	0.10	0.16	0.26
F. Preparation Cost Range, $\$/10^6$ Btu	0.07-0.18	0.07-0.18	0.07-0.18	0.07-0.18	0.07-0.18	0.07-0.18
G. Raw Coal Costs, $\$/10^6$ Btu	1.00	1.00	1.00	1.00	1.00	1.00
H. Coal Replacement Costs, $\$10^6$ Btu	0.08	0.08	0.08	0.04	0.04	0.04
I. Total PCC Costs, $\$/10^6$ Btu	0.15-0.26	0.15-0.26	0.15-0.26	0.11-0.22	0.11-0.22	0.11-0.22
J. PCC + FGD Control Costs, lb $SO_2/10^6$ Btu	0.24-0.35	0.33-0.44	0.41-0.52	0.21-0.32	0.27-0.38	0.38-0.48
IV. Required Coal Cleaning Cost Benefits to Break Even[b], $\$/10^6$ Btu	(0.04)-0.07(-0.01)-0.10		0.04-0.15	0-0.12	0.05-0.16	0.11-0.21

(a) Cleaned at 3/8 inch top size and 1.6 s.g.
(b) Line III-J Costs - Line II-B Costs

Table 8. (continued)

	Emission Control Level 1	Emission Control Level 2	Emission Control Level 3	Emission Control Level 1	Emission Control Level 2
I. A. Coal Region	Eastern Midwest			Western	
B. Raw Coal Sulfur Level, lb $SO_2/10^6$ Btu	6.5	6.5	6.5	1.1	1.1
C. Allowable Emission Limit, lb $SO_2/10^6$ Btu	2.5	1.2	0.65	0.5	0.11
D. Sulfur Removal Required, %	38.5	81.5	90.0	54.5	90.0
II. FGD Control Costs					
A. FGD Sulfur Removal Requirements, lb $SO_2/10^6$ Btu	4.0	5.3	5.85	0.6	0.99
B. Cost of 500 MW at 90 % Removal Efficiency, $\$/10^6$ Btu	0.35	0.40	0.43	0.22	0.230
III. PCC + FGD Control Costs					
A. Sulfur Removal by PCC(a) lb $SO_2/10^6$ Btu	2.3	2.3	2.3	0.2	0.2
B. Sulfur To Be Removed by FGD, lb $SO_2/10^6$ Btu	1.7	3.0	3.55	0.4	0.79
C. FGD Sulfur Removal Requirement, %	40.5	71.4	84.5	44.4	87.8
D. Equivalent FHD Size at 90% Efficiency, MW	225	397	469	245	488
E. FGD Costs at Reduced Size and SO_2 Removal Requirements, $\$/11^6$ Btu	0.14	0.27	0.31	0.13	0.22
F. Preparation Cost Range, $\$/10^6$ Btu	0.07-0.18	.07-0.18	0.7-0.18	.07-0.18	.07-0.18
G. Raw Coal Costs, $\$/10^6$ Btu	1.00	1.00	1.00	0.65	0.65
H. Coal Replacement Costs, $\$/10^6$ Btu	0.05	0.05	.05	.02	.02
I. Total PCC Costs, $\$/10^6$ Btu	0.12-0.23	0.12-0.23	0.12-0.23	0.09-0.20	0.09-0.20
J. PCC + FGD Control Costs, lb $SO_2/10^6$ Btu	0.26-0.37	0.39-0.50	0.43-0.54	0.22-0.33	0.31-0.42
IV. Required Coal Cleaning Cost Benefits to Break Even(b), $\$/10^6$ Btu	(-0.09)-0.02	(-0.1)-0.10	0-0.11	0-0.11	0.08-0.19

(a) Cleaned at 3/8 inch top size and 1.6 s.g.
(b) Line III-J Costs - Line II-B Costs

A range of coal preparation costs was assumed. The lower costs are applicable to coals which could be easily cleaned in a relatively simple plant configuration. The higher costs correspond to the higher levels of cleaning which are required for increased sulfur removal and Btu recovery. Sulfur removal and Btu enhancement equivalent to crushing 3/8 inch top size and separation at 1.6 specific gravity were assumed for all cases. Coal replacement costs were calculated for each of the coals assuming Btu recoveries equivalent to the average coal for each region. Coal replacement costs were assumed to be independent of the plant operating and maintenance costs; i.e., they were held constant over the range of preparation costs.

The results of the analyses were expressed in terms of the non-FGD coal cleaning cost benefits which would be required to make PCC + FGD cost competitive with FGD; i.e.,

$$\begin{array}{c}\text{Required Coal Cleaning}\\\text{Cost Benefits}\end{array} \geq \text{Cost (PCC + FGD) -Cost (FGD)}$$

For some conditions the use of cleaned coals reduces FGD costs to the point where the reduced FGD costs are equal to the coal cleaning costs. This is the "break-even" point at which no other (non-FGD) cost benefits are needed to make the costs of both options equal.

FGD cost offsets from using cleaned coal result from two factors: a reduction in the volume of flue gases treated and a reduction in the sulfur which must be removed from the treated volume. Coal cleaning can reduce the amount of sulfur which must be removed and can at the same time provide for allowable bypass to reduce the volume treated.

For high sulfur coals the break-even point is approached as the percentage of bypass is increased. For low sulfur coals the potential sulfur reduction savings may not be sufficient to offset the cost of cleaning, even under high bypass conditions. Alternatively, regulations which require sulfur removal requirements in excess of 90 percent may not allow for enough bypass (volume) cost reductions to offset the costs of cleaning.

The most likely candidates for PCC + FGD are those applications which use high sulfur coals and which do not require total sulfur removals greater than 90 percent. If the revised NSPS require 90 percent sulfur removal and do not specify an emission floor, then PCC + FGD would not be competitive with FGD unless there are substantial non-FGD cost benefits associated with the use of cleaned coal.

Non-FGD cost benefits may range from $1.00/ton to $6.00/ton (approximately $\$0.04/10^6$ Btu to $\$0.24/10^6$ Btu). If actual cost benefits are at the high end of this cost range, then PCC + FGD may be the most cost effective method of complying with SO_2 emission regulations. If the cost benefits are at the lower end of this cost range, it may not be cost effective to clean coals prior to scrubbing.

CONCLUSIONS

Physical coal cleaning is now used in a limited number of cases to remove sulfur for compliance with SIP SO_2 emission regulations. Tightening and strict enforcement of the SIP regulations may increase the demand for desulfurization of high sulfur Midwestern and Eastern coals by cleaning.

The demand for cleaned coals for compliance with current NSPS for utility boilers is not expected to increase. Boilers subject to this regulation have already selected a control method consisting of low sulfur coal, coal cleaning, or FGD. All new utility boilers will be subject to the revised NSPS.

The percentage reduction specifications of the revised NSPS for utility boilers will essentially preclude the use of coal cleaning as a sole method for complying with the SO_2 control reouirements of these regulations. Combinations of coal cleaning and FGD as a compliance technique will only be used where the combined control approach is more cost effective than FGD alone or where FGD cannot achieve the emission requirements because of an unusually high coal sulfur content. It is also possible that combinations of coal cleaning and FGD will be required to achieve LAER requirements in non-attainment areas or BACT requirements in clean air areas.

It is probable that industrial boiler NSPS which are now being considered by EPA will permit the use of cleaned coals as an SO_2 emission control method in small boilers. Large industrial boilers may find it cost effective to use combinations of coal cleaning and FGD for compliance with SO_2 emission regulations.

The use of PCC + FGD will be the most cost effective method of complying with emission regulations if the reduction in FGD costs and cost benefits not related to SO_2 emission control are greater than the costs of cleaning.

Reductions in FGD costs by PCC can result from a reduction in the volume of flue gas treated or the amount of sulfur removed from the flue gas stream. Reductions in fuel sulfur variability by PCC can lower design safety margins needed to ensure compliance for all fuel sulfur values.

Utility boilers which use high sulfur coals and which require sulfur removals less than 90 percent are likely candidates for PCC + FGD. If the revised NSPS for utility boilers require 90 percent sulfur removal and do not specify an emission floor, then PCC + FGD may not be competitive with FGD unless there are substantial non-FGD cost benefits associated with cleaning.

The range of applications for PCC + FGD in small non-base-loaded utility boilers and industrial boilers may be different from those cited for base-loaded utility boilers. The differentials between PCC and FGD costs for these smaller units may result in different optimal solutions for the range of alternative permissible control strategies.

CONVERSION FACTORS

ton = 0.907 metric tons
lb = 0.436 kg
Btu = 1055.6 Joule
Btu/lb = 2326 Joule/kg
in. = 2.54 cm
°C = 5/9 x (°F-32)
lb./in. = 0.07 kg/cm^2
lb $SO_2/10^6$ Btu = 430 ng SO_2/Joule

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THE INTERAGENCY FLUE GAS DESULFURIZATION EVALUATION STUDY

James C. Dickerman
Radian Corporation
Durham, North Carolina

and

Richard D. Stern
U.S. Environmental Protection Agency
Industrial Environmental Research Laboratory
Research Triangle Park, North Carolina

For Presentation at the Fifth
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THE INTERAGENCY FLUE GAS DESULFURIZATION EVALUATION STUDY

President Carter's National Energy Plan of April 1977 has placed increasing emphasis on United States industries and utilities to convert from oil and natural gas to coal-based energy systems. As coal utilization increases, the potential impact of SO_2 emissions from coal-fired units on air quality will become more significant. Unless adequate steps are taken to control these SO_2 emissions, significant degradation in air quality could result.

This paper presents the results of a study that was conducted pursuant to the President's April 1977 National Energy Plan. The important concerns which led to the study upon which this paper is based can be summarized as follows. The National Energy Policy mandates acceleration of coal usage in the United States. Maintenance of air quality within the context of this increase in coal utilization will require SO_2 emission controls. FGD appears to be the most viable method of SO_2 emission control in the near term (1985). President Carter, in his National Energy Plan, stipulated that the Government would undertake a 6 month study to determine if additional Federal funding would accelerate the commercialization and acceptance of FGD technology. Therefore, this study was conducted to determine if additional Research, Development and Demonstration (RD&D) efforts would accelerate the acceptance of FGD technology. Specific RD&D options and funding levels were also to be identified as an output of this study.

This paper summarizes the procedures used to evaluate the need for increased Federal expenditures to accelerate the use of FGD technology and highlights the findings of that study. First, the project bases and organization are discussed. Next, the approach and methodology used to achieve the study objectives are presented. Finally, the study conclusions, recommendations, and current status of the evaluation are presented.

THE INTERAGENCY FLUE GAS DESULFURIZATION EVALUATION STUDY

INTRODUCTION

This paper is based on a draft report that was completed November 1977 pursuant to the President's National Energy Plan which mandated conversion from an oil- and natural gas-based energy system to one based on coal. The President requested this study to identify if and what government support is necessary to accelerate the development and commercialization of flue gas desulfurization technology in order to permit expanded use of coal without environmental degradation. Although the draft report was completed over a year ago, programs, schedules, and funding levels may be out of date; however, its conclusions concerning flue gas desulfurization (FGD) research opportunities are still valid. In fact, several research opportunities identified in the study are currently being implemented.

As coal usage increases across the United States, the potential impact of SO_2 emissions from coal-fired units on air quality will become more significant. Unless adequate steps are taken to control these SO_2 emissions, significant degradation in air quality could result. There are many alternatives for reducing SO_2 emissions from coal utilization. These include coal cleaning, coal liquefaction, coal gasification, fluidized bed combustion, and FGD. Of these alternatives, only coal cleaning and FGD are currently being commercially applied. It now appears that competing SO_2 emission control technologies will not see widespread applicability before 1985. Since coal cleaning is not attractive for all coals, FGD must be considered to be the most promising near-term alternative for controlling SO_2 emissions across the United States.

This study to determine if additional Federal funding would accelerate the commercialization and acceptance of FGD technology, utilized the following approach.

1) A review of existing and proposed FGD technologies to identify the need for additional research, development, and demonstration (RD&D) activities.

2) An identification of specific RD&D options.

3) An estimate of the level of funding necessary to complete the recommended RD&D objectives.

In particular, advanced FGD concepts were reviewed to determine if they offered the potential for enhancing or improving FGD technologies that are commercial or are nearing commercialization, or if they offered the potential for new and improved approaches to FGD.

GROUND RULES AND ASSUMPTIONS

The following ground rules and assumptions were used for this study:

1) The study would be completed in 6 months.

2) There would be extensive coordination and review within the Federal government.

3) Both processes and subsystems would be evaluated. Processes are defined as complete FGD systems; subsystems are those parts of FGD processes that could be used virtually interchangeably in several FGD processes. For example, within this definition, lime/limestone scrubbing is considered to be a process. The Allied Chemical SO_2 reduction system used with the Wellman Lord installation at NIPSCO is a subsystem because it could be used with any regenerable FGD process that produces a concentrated SO_2 stream.

4) Processes and subsystems that have demonstrated and/or have capability for nitrogen oxides (NO_x) or fine particle removal would be given special consideration.

5) Processes and subsystems recommended for additional funding would have to show either economic, environmental, or technological advantages over existing or developing FGD systems.

6) Processes and subsystems evaluated would have to be capable of being commercialized in a time frame competitive with alternative technologies (by 1985).

The 6 month time frame for the study was necessary to provide timely inputs to EPA's FY78 and FY79 program plans. Extensive coordination and review within the Federal Government was desired in order to ensure maximum objectivity and provide a broad perspective for the study.

The decision to evaluate both processes and subsystems was made to provide maximum flexibility. By evaluating subsystems, it was possible to identify potential improvements to existing and developing processes as well as entirely new process approaches. Processes/subsystems that had the potential for NO_x or fine particle removal were favored because of the obvious advantage they would offer in efforts to expand coal use without environmental degradation.

The requirements that processes/subsystems must be capable of being commercialized in a time frame competitive with alternative technologies and must show either economic, environmental, or technological advantages over existing or developing FGD systems were consistent with accelerating the rate of application of the technology. Any process/subsystem development that improves the reliability, lowers the cost, or reduces the secondary environmental impact of FGD technology could accelerate the rate of application of the technology. Likewise, since FGD is the most promising means of SO_2 emission control until alternatives (such as FBC) are developed, process improvements in FGD technology are particularly important in the period between the present and the time that alternative technologies are commercialized. The elimination of certain processes/subsystems from consideration in this study, however, does not imply that such processes/subsystems will not be worthy of further evaluation or development at a later time.

ORGANIZATION

This study was directed by Richard D. Stern, EPA, Industrial Environmental Research Laboratory - Research Triangle Park (IERL-RTP). To ensure maximum objectivity and incorporate a broad perspective, the study utilized three Interagency groups: a Technical Working Group, an Interagency Steering Panel, and a Liaison Group.

Technical Working Group

This group was chaired by Stephen J. Gage, then Acting Assistant Administrator for Research and Development, EPA. Other members of the group were: Lewis Faucett, Tennessee Valley Authority (TVA); Stuart Dalton, Electric Power Research Institute (EPRI); Myron Gottlieb, Department of Energy (DOE); Lawrence H. Weiss, Chem Systems; and A. V. Slack and Milton R. Beychok, independent consultants. The functions of the Technical Working Group were:

1) To assist in the development of screening criteria to be applied to a comprehensive list of FGD processes/subsystems.
2) To review the initial screening results.
3) To assist in the development of evaluation criteria to be applied to the processes/subsystems which were selected for detailed evaluation.
4) To conduct the detailed evaluation of candidate processes/subsystems.
5) To assess the schedule and costs of the process/subsystem RD&D options.
6) To make recommendations regarding RD&D opportunities.
7) To review the draft final report.

Interagency Steering Panel

The Interagency Steering Panel provided overall guidance in approach and methodology and reviewed study products with emphasis on the final report. Members of the Interagency Steering Panel were: Samuel Biondo, DOE: Bernard Chew, DOE; Howard Feibus, DOE; Stephen J. Gage, EPA; S. William Gouse, DOE; G. R. Hall, DOE; Gerald Hollinden, TVA; David Israel, DOE;

K. H. Jones, Council of Environmental Quality (CEQ); Judy Kammins, Office of Management and Budget (OMB); Rafael Kasper, Office of Science and Technology (OST); Susan Hickey, Federal Energy Administration (FEA); Richard Hertzberg, OMB; C. Morgan Kinghorn, OMB; William E. Mott, DOE; Frank T. Princiotta, EPA; Jack Silvey, DOE; and Kenneth Woodcock, DOE.

Liaison Group

The Liaison Group provided guidance and comment from other organizations which were felt to have interest in the study. Liaison with these organizations was utilized to obtain the broadest possible perspective for the study. Included in the Group were: R. W. Crozier, National Academy of Engineering (NAE); Lloyd Taylor, Science Advisory Board (SAB); H. W. Elder, TVA; and Kurt Yeager, EPRI. Interface with the Liaison Group was handled by Richard D. Stern of EPA's IERL-RTP.

APPROACH AND METHODOLOGY

The approach developed for completing the program objectives is shown schematically in Figure 1. Shortly after initiation of the study, EPA contractors, with the aid of consultants, developed an initial list of processes/subsystems and an initial set of evaluation criteria. A public meeting was held in Washington to acquaint the public with the scope and objectives of the study and to solicit public comment. Results of this meeting added two processes to the comprehensive list (see Table 1) for evaluation.

A series of meetings was then held with the Technical Working Group, contractors, Steering Panel, and the Liaison Group to review and comment on the initial process/subsystem list shown in Table 1 and initial screening criteria shown in Table 2. Once these were approved, technical and economic data were gathered on each process/subsystem and a series of Information Survey Papers was prepared.

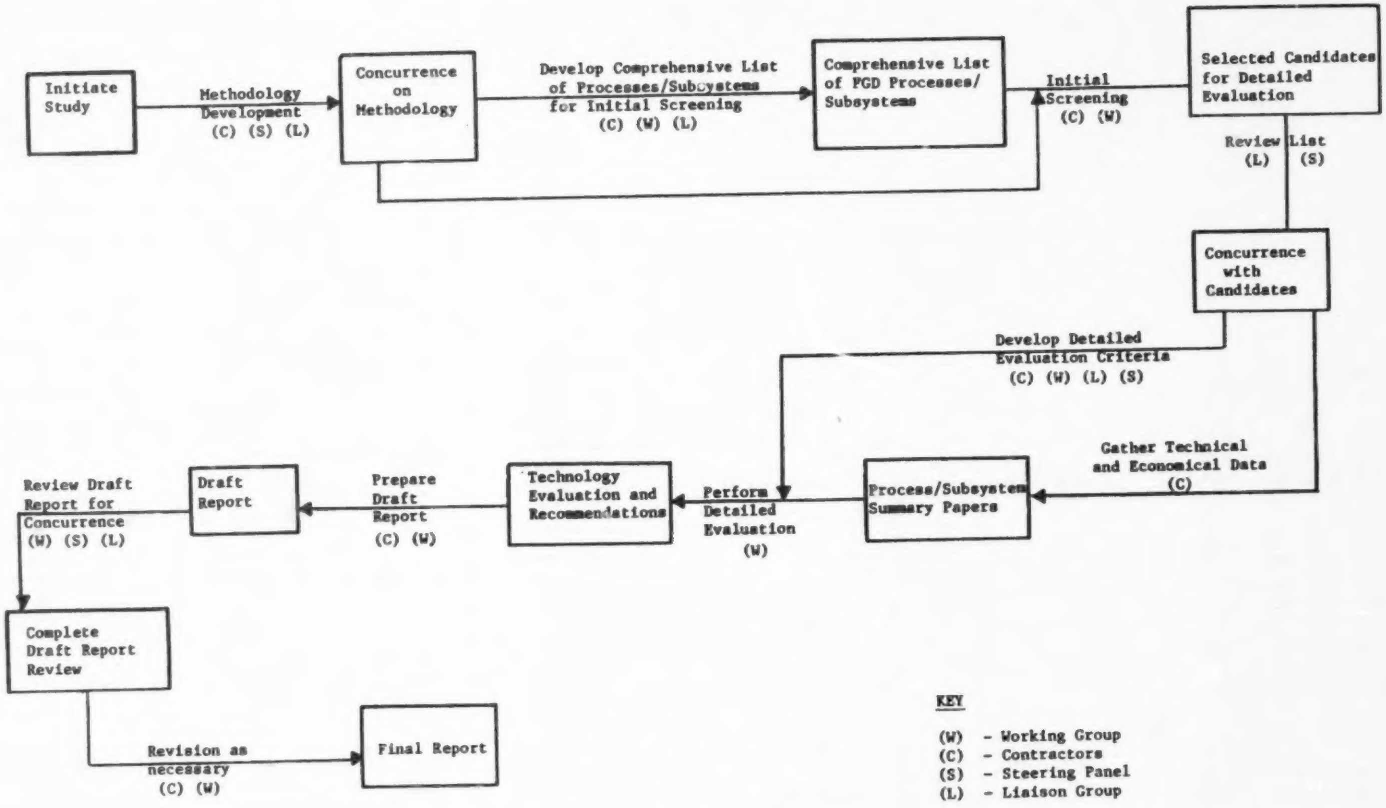
Initiate Study
Methodology Development (C) (S) (L)
Concurrence on Methodology
Develop Comprehensive List of Processes/Subsystems for Initial Screening (C) (W) (L)
Comprehensive List of FGD Processes/ Subsystems
Initial Screening (C) (W)
Selected Candidates for Detailed Evaluation
Review List (L) (S)
Concurrence with Candidates
Develop Detailed Evaluation Criteria (C) (W) (L) (S)
Gather Technical and Economical Data (C)
Process/Subsystem Summary Papers
Perform Detailed Evaluation (W)
Technology Evaluation and Recommendations
Prepare Draft Report (C) (W)
Draft Report
Review Draft Report for Concurrence (W) (S) (L)
Complete Draft Report Review
Revision as necessary (C) (W)
Final Report
KEY
(W) - Working Group
(C) - Contractors
(S) - Steering Panel
(L) - Liaison Group

Figure 1. Interagency FGD Technology Evaluation-Study Approach and Methodology.

TABLE 1. COMPREHENSIVE LIST OF POTENTIAL CANDIDATES FOR DETAILED EVALUATION

Throwaway Systems	Wet Regenerable Systems	Dry Regenerable Systems	Subsystems	Others
Agglomerating Cone	Ammonia Bisulfate (ABS)	Alkalized Alumina	Allied Reduction Process	Allied Chemical Electrodialysis
Alkaline Ash Scrubbing	Ammonia Evaporative Crystallization	Bergbau-Forschung	ASARCO Sulfur Plant	Allied Chemical Membrane Process
Ammonia Scrubbing/H_2SO_4 Acidulation (Cominco, Ammsox)	Ammonia/IFP (Catalytic, Research Cottrell)	Bureau of Mines MnO_2	Calcium Sulfate Regeneration	Barium Carbonate
Ammonia Solution Double Alkali	Aquaclaus	Carbon Adsorption/Inert Gas Stripping (Reinluft/Chemiebau, Sumitomo)	Citrate Double Loop Regeneration	Battelle Fused Salt
Ammonia Vapor Double Alkali (Ugine Kuhlman, Nippon Kokan)	ASARCO Dimethylaniline	Copper Oxide Adsorption (Shell, Esso B&W, Bureau of Mines)	Claus Process	Boliden
Asahi Chemical	Consol	Callery Chemical	Cocurrent Scrubbing	Catalox
$C\ell O_2$	Grillo	Hormel	Dilute System Sulfate Removal	Cooper
Calcium Chloride (Hölter, Kobe)	ICI Steam Stripping	Houdry	Direct Production of Sulfur from $MgSO_3$	Condensation and Reaction with Fly Ash
Calsox	Johnstone Zinc Oxide	In-Situ CO Reduction (Chevron)	Forced Oxidation of $CaSO_3$	Dry Removal with Ground Lime
Carbon Adsorption (Hitachi, Lurgi-Sulfacid)	Lurie Sodium Aluminate	In-Situ H_2S Reduction (Peter Spence, Princeton Chemical, Ontario Research Foundation)	IFP Subsystems	Dry Sorbents
Chisso Engineering	Maget	Integrated Cat-Ox	Ionics Electrolytic Regeneration	Electrochemical Concentration
Chiyoda Thoroughbred 101	McKee	Kiyoura-TIT	Mass Transfer Additives with Lime/Limestone Scrubbing	Esso (V_2O_5)
Chiyoda Thoroughbred 102	Melamine	Mitsubishi Manganese Oxide	Production of Reducing Gas	Goodrich
Chiyoda Thoroughbred 121	Mitsui Eng. & Shipbuilding	Purasiv S	RESOX	ITT
Double Alkali with Limestone (Showa-Denko, Kureha-Kawasaki)	Molten Carbonate	Rohm & Haas Resin Adsorption	Sludge Stabilization	McGauley
Dowa Aluminum Sulfate	Molten Potassium Carbonate/Thiocynate	Shuffman	Tampella Recovery Process	MgO Based Double Alkali
Dry Alkalis plus Fabric Filter	NIIOGAZ Ammonia/Steam Stripping	Sumitomo Heavy Industries		Reduction with "Blue Gas"
Ebara-Jaeri	NIIOGAZ Magnesium Oxide	TOPSOE		Siemens
Ishikawajima-Harima HI	NOSOX	Tyco Chamber		Soil Process
IPRAN	Peabody	Unitika		Sulfuric and Nitric Acid Recovery Method
Kawasaki MgO/Lime	Potassium Bisulfite	Uranium Oxide		
Koyo	Potassium Formate	Westvaco		
Krebbs-Neville	Potassium Sulfite			
Kurabo	Ralph Parsons			
Kureha (SO_2/NO_x)	Spring-Nobel Hoechst			
Kureha Sodium Acetate	Stone & Webster/Ionics			
Lead-Zinc Ore	Sulf-X			
Lewis Process	TSK Sulfix			
Micropul	UOP Sulfoxel			
Mitsubishi Heavy Industries				
Moretana Calcium				
Moretana Sodium				
Red Mud				
RIIC				
SCRA				
Sea Water				
Simon Carves				
Sodium Carbonate Scrubbing				
Sodium Hypochlorite				
Sulfurtain				

TABLE 2. INITIAL SCREENING EVALUATION CRITERIA FOR CANDIDATE PROCESSES/SUBSYSTEMS

I. Minimum Requirements	Go/No Go
Processes/subsystems which fail to meet any of these requirements will not be considered further	
A. A process/subsystem must be able to achieve compliance with the minimum assumed future NSPS for SO_2	
B. A process/subsystem must be capable of being demonstrated by 1985	
C. A process/subsystem must be applicable to treating flue gases from combustion of coal	
D. Adequate information must be available to enable process/subsystem evaluation	
II. Environmental and Energy Considerations	30 points
A. Compliance with SO_2 regulations (Assumes 0.2 kg $SO_2/10^9$ Joules; 0.4 lb $SO_2/10^6$ Btu)	(6 points)
6 - potential to comply with future regulations with high sulfur coals (96%)[a]	
4 - potential to comply with future regulations with medium sulfur coals (94%)	
2 - potential to comply with future regulations with low sulfur coals (80%)	
B. Potential for multipollutant removal	(3 points)
3 - more than one	
1 - one	
0 - none	
C. Performance growth potential	(3 points)
D. Secondary pollutant problems	(3 points)
3 - none	
2 - minor	
0 - major	
E. Relative Energy Requirements Since material and energy balances will not be available for many of these processes, the following qualitative rating factors will be used:	

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[a] SO_2 Removal

TABLE 2. (Continued)

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1) Reheat (3 points)
   - 3 - none
   - 2 - possibly
   - 1 - from saturation (52°C-79°C; 125°F - 175°F)

2) Reducing Gas (3 points)
   - 3 - none
   - 1 - CO and H_2
   - 0 - H_2 only

3) L/G Ratio, Pressure Drop (3 points)
   - 3 - low
   - 1 - medium
   - 0 - high

4) Other Energy (3 points)
   - 3 - low
   - 1 - medium
   - 0 - high

F. Raw Material Requirements (3 points)

   - 3 - uses less than 2 that are readily available
   - 2 - uses less than 2 that are not readily available
   - 0 - uses more than 2 that may not be readily available

III. Development Status 25 points

A. State of Development (10 points)

   - 10 - demonstration (>30 MW) coal-fired
   - 8 - demonstration (>30 MW) oil-fired
   - 7 - prototype (10-30 MW) coal-fired
   - 5 - prototype (10-30 MW) oil-fired
   - 4 - pilot (1-10 MW) coal-fired
   - 3 - pilot (1-10 MW) oil-fired
   - 2 - bench scale
   - 1 - conceptual

B. Degree of Integration (5 points)

   - 5 - totally integrated
   - 3 - sorption and regeneration integrated
   - 0 - nothing integrated

C. Use of Currently Commercialized Technology (5 points)

   - 5 - in all process sections
   - 3 - in at least two process sections
   - 2 - in one process section
   - 0 - not at all

TABLE 2. (Continued)

	D. Scale-up Problems	(5 points)
	5 - no apparent problems	
	2 - potential problems with complexity or scale-up of a new or unique processing step	
	0 - potential scale-up problems in several areas	
IV.	Economic and Technological Considerations (applied to major sections and subsystems) Capital requirements may be estimated by considering:	25 points
	A. Relative Chemical Complexity	(15 points)
	15 - simple, workable	
	10 - complex, workable	
	5 - simple, questionable	
	0 - complex, questionable	
	B. Relative Mechanical Complexity	(10 points)
	10 - simple	
	5 - moderately complex	
	0 - very complex	
V.	Applicability	20 points
	A. Suitability for Utility Applications	(10 points)
	1) Separability of Process Steps	(3 points)
	3 - can be easily decoupled	
	1 - difficult decoupling	
	0 - decoupling not practical	
	2) Load Following Capability	(3 points)
	3 - follows load changes rapidly	
	1 - follows load changes slowly	
	0 - process sections do not readily follow load changes	
	3) Retrofitability	(2 points)
	2 - easy to retrofit	
	1 - will require minor modifications	
	0 - significant problems	
	4) By-product Utilization/Marketability	(2 points)
	2 - readily marketable	
	1 - marginally marketable	
	0 - unmarketable	

TABLE 2. (Continued)

B. Suitability for Industrial Application	10 points
1) By-product Utilization/Marketability 3 - readily marketable 1 - marginally marketable 0 - unmarketable	(3 points)
2) Ability to Modularize the Process 3 - suitable for small (>10 MW) modules) 1 - suitable for medium (10-40 MW) modules 0 - suitable for large (>40 MW) modules	(3 points)
3) Separability of Process Steps 2 - can be easily decoupled 1 - difficult decoupling 0 - decoupling not practical	(2 points)
4) Retrofitability 2 - easy to retrofit 1 - will require minor modifications 0 - significant problems	(2 points)

Next, the initial screening criteria were applied to the comprehensive process/subsystem list to select those processes that offered the most potential for accelerating the commercialization and application of FGD technology. These selected, or candidate, process were then evaluated in more detail in the next phase of the project to identify specific RD&D opportunities. This screening was necessary to reduce the task of detailed evaluation to a reasonable size by eliminating those processes that appeared to have the least potential for near term improvement of FGD technology. The elimination of some processes from consideration in this study does not imply that such processes will not be worthy of development or further evaluation at a later date. The initial screening focused on the selection of candidate processes in two main areas: 1) those that enhance or improve technologies currently used in commercial and developing processes, and 2) those that, if developed, would offer new and improved approaches to FGD. Commercial processes, as well as processes being developed under Federal funding, formed a basis for comparison with the processes/subsystems being screened. For purposes of this evaluation, the following were considered to be currently commercial processes or processes being developed under Federal funding:

Lime/Limestone Wet Scrubbing
Magnesia Slurry Scrubbing
Wellman-Lord
Sodium/Lime Dual Alkali
Citrate Buffered Absorption
Rockwell International's Aqueous Carbonate Process

A total of 138 processes were screened to examine their potential for offering technological and/or economic advantages over commercialized and developing FGD processes. Results of the screening shown in Table 3 indicated that 13 processes/subsystems should be evaluated in detail, along with the 6 commercial and developing processes listed above, for the purpose of identifying RD&D opportunities that would accelerate the commercialization of FGD.

TABLE 3. PROCESSES SELECTED FOR FURTHER EVALUATION

Throwaway	Wet Regenerable	Dry Regenerable	Subsystems
Calcium Chloride	Ammonia/IFP	Bergbau-Forschung	RESOX
Chiyoda Thoroughbred 121	[a]Atomics International Aqueous Carbonate Process (ACP)	Copper Oxide (Shell-UOP)	
Dowa Aluminum Sulfate	[a]Citrate	Integrated Cat-Ox	
Kurabo	[a]Magnesia Slurry	Sorption/Steam Stripping (Rohm & Haas Resin Adsorption Process)	
[a]Lime/Limestone	Melamine		
Sodium/Lime Double Alkali	Sorption/Steam Stripping		
Sodium Throwaway Systems	[a]Wellman-Lord		

[a] Commercial or current federally funded processes

To assist the Technical Working Group in evaluating and comparing the processes in a consistent impartial manner, a set of evaluation criteria was developed for application to each of the 19 processes selected as potential candidates for RD&D funding. In addition to providing a means of ranking the processes, a consistent set of evaluation criteria was judged to be necessary to evaluate RD&D options for the following reasons:

- Evaluation criteria formalize the evaluation procedure and allow the evaluators to compare their process evaluations on a consistent basis. Any discrepancies in the evaluation can then be isolated and discussed.
- Specific evaluation criteria force analysis of all process advantages and disadvantages.
- Key process deficiencies can be identified through the use of evaluation criteria.

The criteria developed for the detailed process evaluations are presented in Table 4.

The objectives of the detailed process evaluations were to identify specific RD&D programs and to identify specific benefits to be derived from the successful implementation of these programs that would achieve the overall goal of accelerating the commercialization of FGD technology. A list of RD&D recommendations was developed based on the detailed evaluations, and a draft report was prepared and submitted for review.

CONCLUSIONS AND RECOMMENDATIONS

To enhance the utilization of coal and to minimize its environmental impact, the Technical Working Group developed the following conclusions:

TABLE 4. DETAILED EVALUATION FOR PROCESS/SUBSYSTEMS SELECTION

I. Environmental and Energy Considerations — 25 points

A. Compliance with SO_2 Regulations — (5 points)

5 - ability to comply with future regulations with high sulfur coals (96%)
3 - ability to comply with future regulations with medium sulfur coals (94%)
1 - ability to comply with future regulations with low sulfur coals (80%)

B. Potential for Multipollutant Control — (3 points)

3 - significant removal of 1 or more
2 - some removal of 1 or more
1 - no removal of other pollutants

C. Secondary Pollution Problems - Air — (2 points)

2 - no air emission problems
1 - potential air emission problems
0 - major air emission problems

D. Secondary Pollution Problems - Liquid — (2 points)

3 - no liquid wastes
2 - liquid wastes that can be easily treated
1 - liquid wastes that require unusual treatment

E. Secondary Pollution Problems - Solid — (3 points)

3 - no solid wastes
2 - solid wastes that can be easily disposed of
1 - solid wastes that require unusual treatment

F. Energy Intensiveness — (10 points)

- Chemical and electrical energy requirements compared with technology currently commercialized or under development
  6-10 - lower energy requirements
  5 - average energy requirements
  1-4 - higher energy requirements

II. Development Status — 15 points

A. Overall Process/Subsystem Development Status — (5 points)

- How long has developer studied and worked his process/subsystem?

- How much of the process/subsystem is technically well known and well founded?

TABLE 4. (Continued)

- At what level has it been operated and for how long?
  - Bench Scale,
  - Laboratory,
  - Pilot Plant,
  - Small Prototype,
  - Paper Study.
- Have all sections been operated as a chemically or mechanically integral system?

B. Process/Subsystem Operations and Use of Commercialized Technology (5 points)

5 - operations in common practice
3 - industrially demonstrated
2 - some uncommon methods involved
0 - several unproven processing steps

C. Process Controllability (5 points)

5 - simple process without many sensitive control requirements
3 - simple process with sensitive control requirements
2 - complex process without many sensitive control requirements
0 - complex process with sensitive control requirements

III. Economic and Technological Considerations 35 points

A. Capital Investment Costs (15 points)

15 - 40% less than median
14 - 30% less than median
12 - 20% less than median
10 - 10% less than median
8 - median
6 - 10% more than median
4 - 20% more than median
2 - 30% more than median
0 - 40% more than median

B. Annualized Costs (15 points)

15 - 40% less than median
14 - 30% less than median
12 - 20% less than median
10 - 10% less than median
8 - median
6 - 10% more than median
4 - 20% more than median
2 - 30% more than median
0 - 40% more than median

TABLE 4. (Continued)

C. Economic Sensitivity to Technological Factors (5 points)

5 - little uncertainty in technological factors affecting cost
3 - some uncertainty in technological factors affecting cost
0 - considerable uncertainty in technological factors affecting cost

IV. Applicability 25 points

A. Suitability for Utility Application (10 points)

- Retrofitability
- By-product Utilization/Marketability
- Separability and Process Steps
- Load Following Capability
- Acceptability of Processing Techniques to the Industry

B. Suitability for Industrial Application (10 points)

- Retrofitability
- By-product Utilization/Marketability
- Separation of Process Steps
- Ability for Process Modularization
- Acceptability of Processing Techniques to Small Boiler Applications

C. Flexibility to Provide Alternate Products (3 points)

3 - can product sulfur or acid
2 - can produce both sulfur and acid with significant modifications
1 - produces either sulfur or acid
0 - produces neither sulfur or acid

D. Regional/Site Considerations (2 points)

2 - process/subsystem suitable over broad range of applications
1 - process/subsystem suitable only for narrow range of applications

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- There are significant potential benefits to be derived from increasing the level of Federal funds for FGD research, development, and demonstration (RD&D) efforts.
- The increased Federal funding should be utilized for research and development projects to improve systems, both throwaway (Lime/Limestone) and regenerable, that have been or are being demonstrated on large-scale equipment.
- Resources should be available for process/subsystem evaluations to assess potential solutions to common FGD problem areas and to continually evaluate new or enhanced developments in FGD technology.

It was the consensus of the Technical Working Group that the highest priority research opportunities were in the area of further development to improve the economics and applicability of systems that have been or are being demonstrated on large-scale equipment. Because of the emphasis upon finding near-term solutions to existing problems, many of the RD&D recommendations were to further investigate process subsystems or process alternatives. These recommendations, in general, concerned RD&D categories that have potential widespread application, especially in wet and throwaway FGD processes. A secondary priority was assigned to processes and process developments that provide longer-term benefits. A discussion of each recommended RD&D opportunity and its expected benefit is presented below.

1) Lime/Limestone Improvements

A) Mass Transfer Additives - An extension of current programs that have focused upon evaluating magnesium as a mass transfer additive for lime/limestone processes was recommended. Specific additives to be examined included both organic and inorganic mass transfer additives such as adipic acid, calcium chloride, and sodium carbonate. The use of a mass transfer additive has the potential for improving the SO_2 removal ability of lime/limestone systems, increasing sorbent utilization, and decreasing process costs. Thus the primary benefits of this research could be an increased level of application of lime/limestone FGD technology and a lower process cost. Initial efforts should involve pilot scale testing followed by demonstration in a prototype unit.

B) Forced Oxidation - EPA has an existing program to evaluate forced oxidation at TVA's Shawnee Steam Plant. Pilot testing of two process variations is currently being conducted. A full-scale demonstration program was recommended for 1979-1980. A full-scale demonstration of this concept could benefit lime/limestone technology in reducing sludge disposal problems, thus removing a major environmental impediment to lime/limestone scrubbing, and increasing its applicability. In addition, a potentially marketable gypsum product may be produced. Since full-scale demonstration could occur by modifying an existing system, this appears to be the next step in commercializing this technology.

C) Contactors - Extension of the current work to evaluate cocurrent flow designs was recommended. The use of a cocurrent scrubber could significantly reduce the cost and complexity of lime/limestone scrubbing. Extension and expansion of current programs appear necessary to conclusively answer operability questions. In addition, testing and demonstration of the Chiyoda Thoroughbred 121 (CT-121) system were recommended, as this system has claimed significant cost reduction over conventional limestone scrubbing due to its simple design.

D) Sludge Disposal - Extension of current sludge disposal activities to evaluate large-scale disposal options (fixation, lining, and ponding) was recommended. Sludge disposal is a major environmental problem in lime/limestone scrubbing. Resolution of this problem on a large scale could enhance the applicability of lime/limestone and dual alkali systems.

E) Hardware Improvements - Problems with mist eliminators, reheaters, and stack liners have been the cause of much of the current FGD system failures. Additional development work to improve these components was recommended as a means of increasing the reliability and operability of wet FGD systems. Demonstration of improved designs for these components on full-scale systems was recommended to gain acceptability by the industry.

2) Sulfur Production with Carbon

A) RESOX - EPRI is funding a RESOX demonstration program in Germany to produce sulfur from the Bergbau-Forschung FGD unit using anthracite coal as the reductant. Evaluation of the RESOX process in a demonstration-sized facility with feed from front-end systems other than Bergbau and evaluation of coals other than anthracite for use as a reductant was recommended. The overall applicability of the RESOX process suffers because it has not been evaluated on other systems. Extension of the EPRI program to test the RESOX process under different conditions could result in the solution of a major problem in FGD: sulfur production without the use of a reducing gas. This could significantly enhance applicability of FGD due to its potential for reducing cost, complexity, and secondary pollution.

B) Rockwell International Regeneration - Extension of the EPRI program to demonstrate and operate the Rockwell International (RI) Regeneration System in an integrated mode to gather data for full-scale design and construction was recommended. RD&D efforts on this system would demonstrate direct conversion of SO_2 to sulfur without a reducing gas plus provide additional design data for the Aqueous Carbonate Process (ACP) to be constructed at Niagara Mohawk under EPA funding.

3) Continuing Evaluation Effort - It was recommended that funds be set aside for use in technical and economic evaluation of specific problems or of processes that in the future may appear to offer advantages over current FGD technology but that were not considered in this study due to their lack of development. This continuing evaluation effort could ensure that future improvements in FGD technology are given a complete and timely evaluation. In addition, it would provide a source of funding for important studies on critical issues in FGD technology that could be of significant benefit in directing future RD&D activities.

4) Magnesia Slurry Scrubbing - Demonstration of sustained sulfuric acid production from a full-scale system would be beneficial in advancing a technology that appears to have potential for very high SO_2 removal. In addition, demonstrating the ability to directly produce sulfur in the calciner will significantly enhance process applicability. Both concepts should be demonstrated and evaluated on a full-scale system.

5) Limestone Use in Sodium Based Dual Alkali Systems - Demonstration of the use of limestone as the regenerant in ongoing pilot work and extension and modification of the EPA sponsored demonstration program at Louisville Gas and Electric (LG&E) to test limestone was recommended. The use of limestone in dual alkali systems has potential for significantly reducing system operating costs because limestone is a less expensive raw material than lime.

6) Sodium System Waste Handling

   A) Disposal - Test of concepts for fixation, plus other methods of sludge handling and disposal from once-through sodium systems, were recommended. The disposal of sodium sludge is a major problem in sodium throwaway processes. A solution to this problem would increase the applicability of this process with potential reductions in both cost and environmental impacts.

B) Regeneration - Evaluation of methods for the regeneration of sodium system by-products and eventual testing was recommended. This program has a similar benefit to the option listed above. It could eliminate disposal problems faced by sodium systems and reduce raw material costs.

7) Application of Low-Btu Gas - Evaluation of the applicability of coal-derived low-Btu gas to regenerate FGD systems was recommended. This would include feasibility studies to evaluate system requirements, followed by demonstration of the applicability of specific gasification systems. The use of coal gasification could eliminate the dependence of some regenerable FGD processes on increasingly costly and scarce supplies of natural gas.

8) Sorption/Steam Stripping - Extension of the EPRI work in the areas of laboratory studies and pilot work, to be followed by demonstration at the 60-100 MW level, was recommended. It was felt that design of a general test facility for the testing of the large number of concepts currently under consideration would be beneficial. The Sorption/Steam Stripping concept appears to have potential for more reliable, lower cost operation than some other regenerable systems and should be assessed.

9) Dowa Test Facility - A pilot plant program to evaluate the feasibility of the Dowa process in coal-fired applications was recommended. This process rated high in the evaluations and appears to have significant advantages over current dual alkali FGD technology in that it uses limestone as a regenerate and produces a marketable quality gypsum by-product. A program of this nature is currently being considered by EPRI and TVA.

10) Integrated Cat-Ox - Additional support for the Cat-Ox process to demonstrate the Integrated Cat-Ox system at the 100 MW level was recommended. The Integrated Cat-Ox process appears to have significant cost advantages over current FGD technology. Successful demonstration of this process could have significant benefits in the areas of cost and process complexity.

11) SULF-X - An evaluation of the SULF-X process to independently analyze vendor claims was recommended. The process has potential as a SO_x/NO_x flue gas treatment process; however, the claims have not been independently evaluated. This evaluation is currently planned by DOE.

The Working Group also recommended that a total level of funding of over $100 million dollars would be necessary to complete the identified RD&D opportunities. It was anticipated that funding of many of the recommended projects would be on a cost-shared basis with the process vendor, host utility or industrial site, EPRI, or another government agency. The estimated total funding level did not attempt to define percentage funding by participant. Figure 2 presents a prioritized listing of the recommended RD&D opportunities along with the then-estimated funding levels and schedules.

SUMMARY

The approach and methodology used in this evaluation have provided an effective means of identifying research opportunities for enhancing the near-term commercialization of FGD technology. Although the recommended program has not yet been formally approved, its impact has been felt in that several of the recommended RD&D opportunities have already been implemented. For example, in the lime/limestone process area, EPA has initiated work to evaluate the effects of adipic acid addition on SO_2 removal, and plans are being made to demonstrate forced oxidation at TVA's Widow's Creek Plant. EPRI has development programs under way to evaluate both cocurrent and sparged

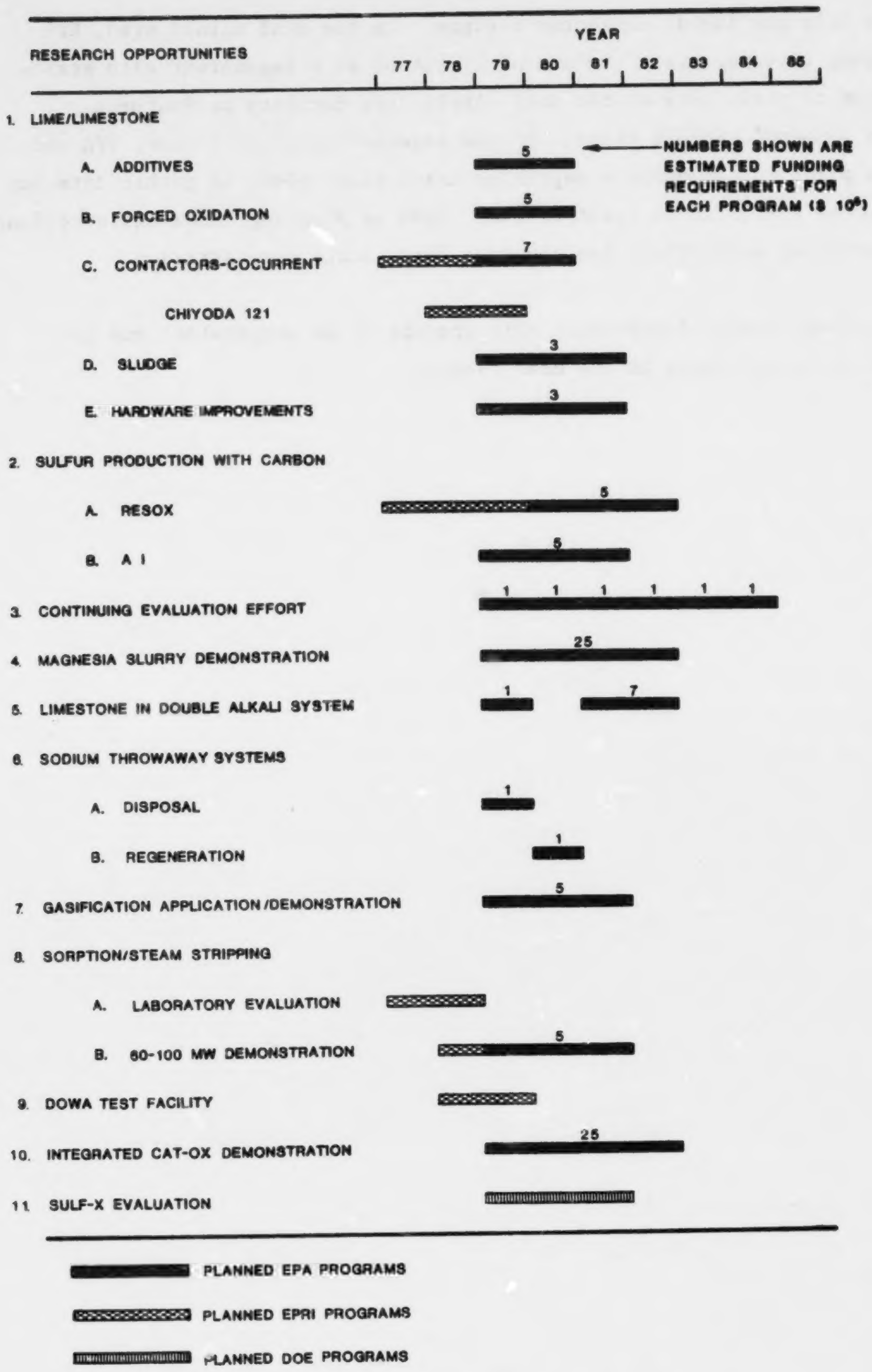

Figure 2. Prioritized RD&D Opportunities.

(Chiyoda 121) gas/liquid contactor designs. In the dual alkali area, EPA is planning to demonstrate the use of limestone as a regenerant with start-up planned in early 1979 at the dual alkali test facility at Southern Services Company's Scholz Plant. In the regenerable process area, TVA and EPA have plans to cosponsor a magnesium oxide pilot plant to gather data for a full-scale system to be built by TVA. EPRI is planning RESOX and sorption/steam stripping evaluations leading to a large-scale demonstration.

The final report documenting this project is in preparation and is expected to be published in the near future.

SESSION 4

UTILITY APPLICATIONS

MICHAEL A. MAXWELL, CO-CHAIRMAN
JULIAN W. JONES, CO-CHAIRMAN

STATUS OF FLUE GAS
DESULFURIZATION
IN THE UNITED STATES

Prepared by

Bernard A. Laseke and Timothy W. Devitt

PEDCo Environmental, Inc.
11499 Chester Road
Cincinnati, Ohio 45246

For Presentation
at the Fifth Symposium on
Flue Gas Desulfurization

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SECTION 1
INTRODUCTION

Pedco Environmental, Inc., under contract to the Environmental Protection Agency (EPA), has closely monitored the growth and use of FGD technology by utilities in the United States and has evaluated FGD technology on both a general and a site-specific basis. The site-specific evaluations are based on visits to plants with operating FGD systems, during which process design and performance information and capital and annual cost data are obtained. A series of reports has been prepared and published on the major operational installations.

Perhaps the most significant product of this project is the periodic summary reports that are issued. These reports provide updated data on the number and capacity of the systems in operation, under construction, or planned and describe the performance of the operating systems during the reporting period. Utility representatives, system suppliers, system designers, regulatory personnel, and others contribute this information voluntarily to facilitate the timely transfer of information in this key technological area. Information provided by utility representatives with operating systems is reported essentially as obtained; little attempt is made to analyze or interpret the data. Information provided by system suppliers and other sources is confirmed with the appropriate utility prior to publication.

The following sections of this paper address some of the highlights of the ongoing survey program.

SECTION 2

OVERVIEW OF FGD TECHNOLOGY

Table 1 lists the total number of FGD installations and their equivalent electrical capacities (in MW) as of the end of November 1978.

TABLE 1. NUMBER AND CAPACITY OF U.S. UTILITY FGD SYSTEMS

Status	Number of units	Capacity, MW
Operational	46	16,054
Under construction	43	17,297
Planned:		
Contract awarded	20	10,690
Letter of intent signed	3	1,960
Requesting/evaluating bids	5	3,100
Considering FGD	27	13,406
Total	144	62,507

As the table shows, 144 FGD systems representing an equivalent electrical capacity of 62,507 MW are in operation, under construction, or planned. Of these systems, 46 are operational (16,054 MW), 43 are under construction (17,297 MW), and 55 are planned (29,156 MW). There are another 55 to 60 plants that will be using FGD systems, but information regarding these systems is not yet ready for public release. These systems will have an equivalent electrical generating capacity between 36,000 and 41,000 MW. To date, 16 systems (1555 MW) have been shut down for various reasons. Three of these systems (425 MW) are continuing to operate, removing primarily fly ash. However, the systems do remove some sulfur dioxide (35 to 50 percent) as a result of alkaline additives put into the scrubbing solution for pH control.

GROWTH TRENDS

Figure 1 illustrates both the number and equivalent capacity of FGD systems as a function of year of start-up. The number of systems requires clarification. A system is defined on the basis of inlet gas ducting configuration. A module or several modules that are commonly ducted to one or more boilers comprise a single system. Thus a single FGD module that treats flue gas from only one boiler is considered a system, just as multiple FGD modules connected through a common duct to multiple boilers are considered one system. On the other hand, when a plant has several boilers ducted to a number of distinct modules or groups of modules without any common ducting between them, that plant is considered to have separate FGD systems.

The values in Figure 1 represent all the FGD systems installed and operated from March 1968 to the end of November 1978, as well as those under construction and planned for installation from December 1978 to 1986. Systems planned for operation beyond 1986 are excluded because they are in the preliminary planning stage and public information is limited.

Figure 2 shows the increase in the projected capacity of FGD systems as a function of the year the estimate was prepared. In November 1974, for example, a total of 37,836 MW of capacity could be identified as in operation, under construction, or

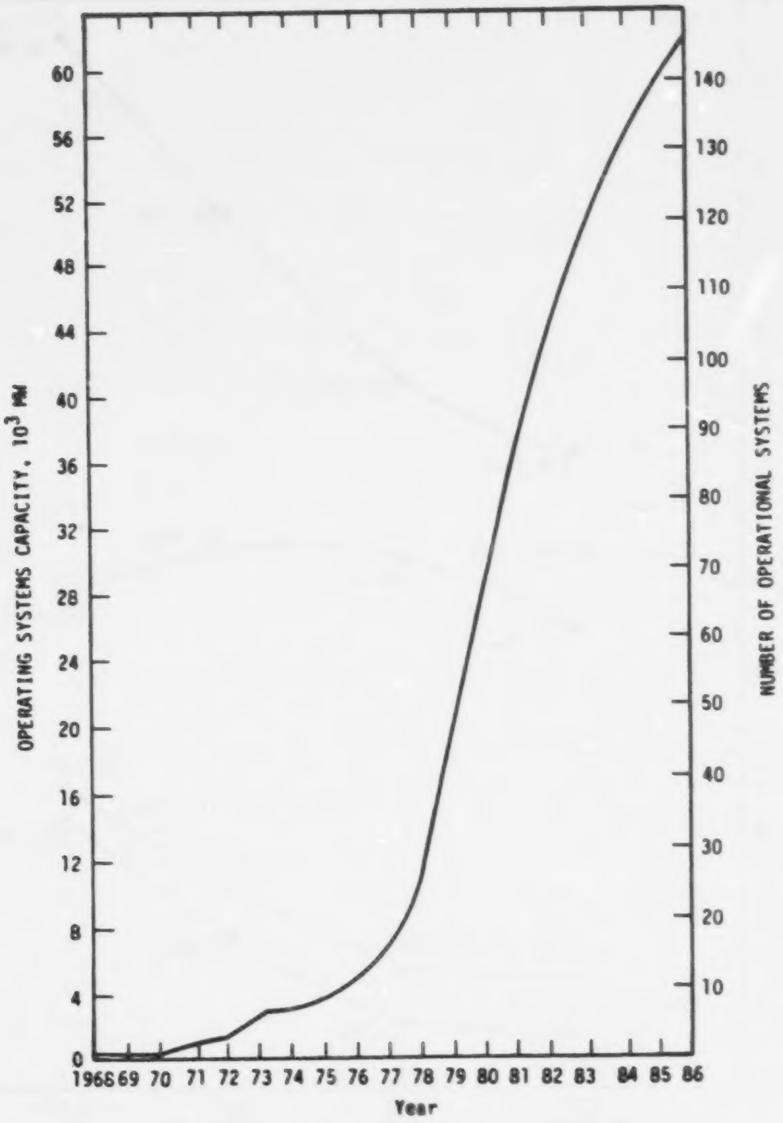

Figure 1. FGD operating capacity through 1986.

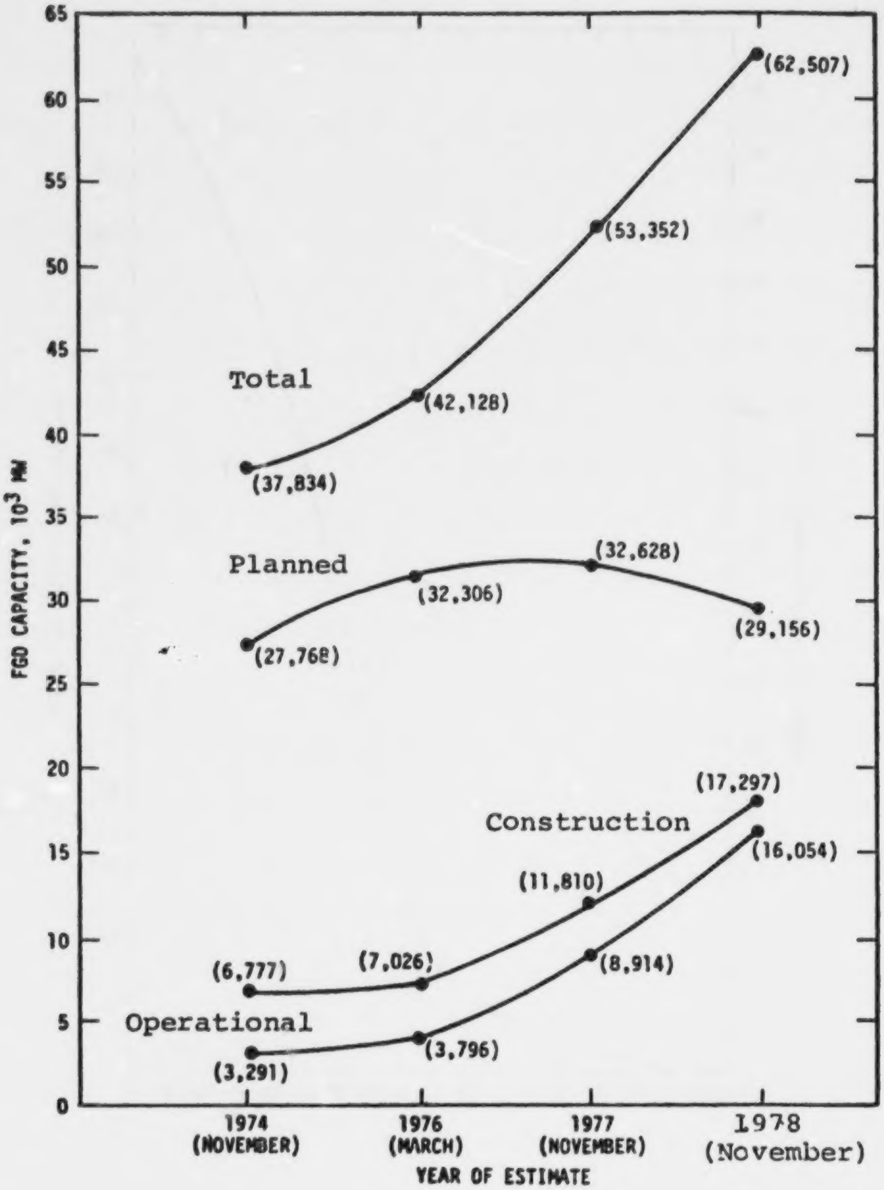

Figure 2. FGD capacity as a function of status and year of estimate

planned; and by the end of November 1978, 62,507 MW is accounted for (this does not include approximately 36,000 to 41,000 MW of planned capacity that cannot be identified at this time).

From 1974 to late 1978 the number of operating systems reported increased from 19 to 46, a 242 percent increase, while the equivalent generating capacity increased from 3,291 MW to 16,054 MW, an increase of 488 percent. The average system size has increased from 173 MW to 434 MW in the same time period, and the capacity associated with full-scale systems has increased from 2,360 MW to 15,882 MW. (Full-scale systems are defined as those that are available for commercial operation on fossil-fuel-fired boilers having a minimum power generating capacity of 100 MW.)

A general uncertainty surrounds projections of the power-generating capacity of new coal-fired boilers. One such projection, developed by PEDCo Environmental from a number of sources (References 1, 2, and 3), is shown in Figure 3. Current coal-fired capacity is indicated as approximately 265 GW, which represents 47 percent of the total power-generating capability of the electric utility industry in the United States. By 1986, this figure is expected to rise to 363.2 GW and to represent 45 percent of the projected total power capacity; and by 1990, it is expected to be 440 GW and to represent 44 percent of the projected total power capacity.

Figure 3 also shows the projected application of FGD systems through 1986. Two major categories of FGD systems are depicted--committed and uncommitted. The committed FGD systems (listed in

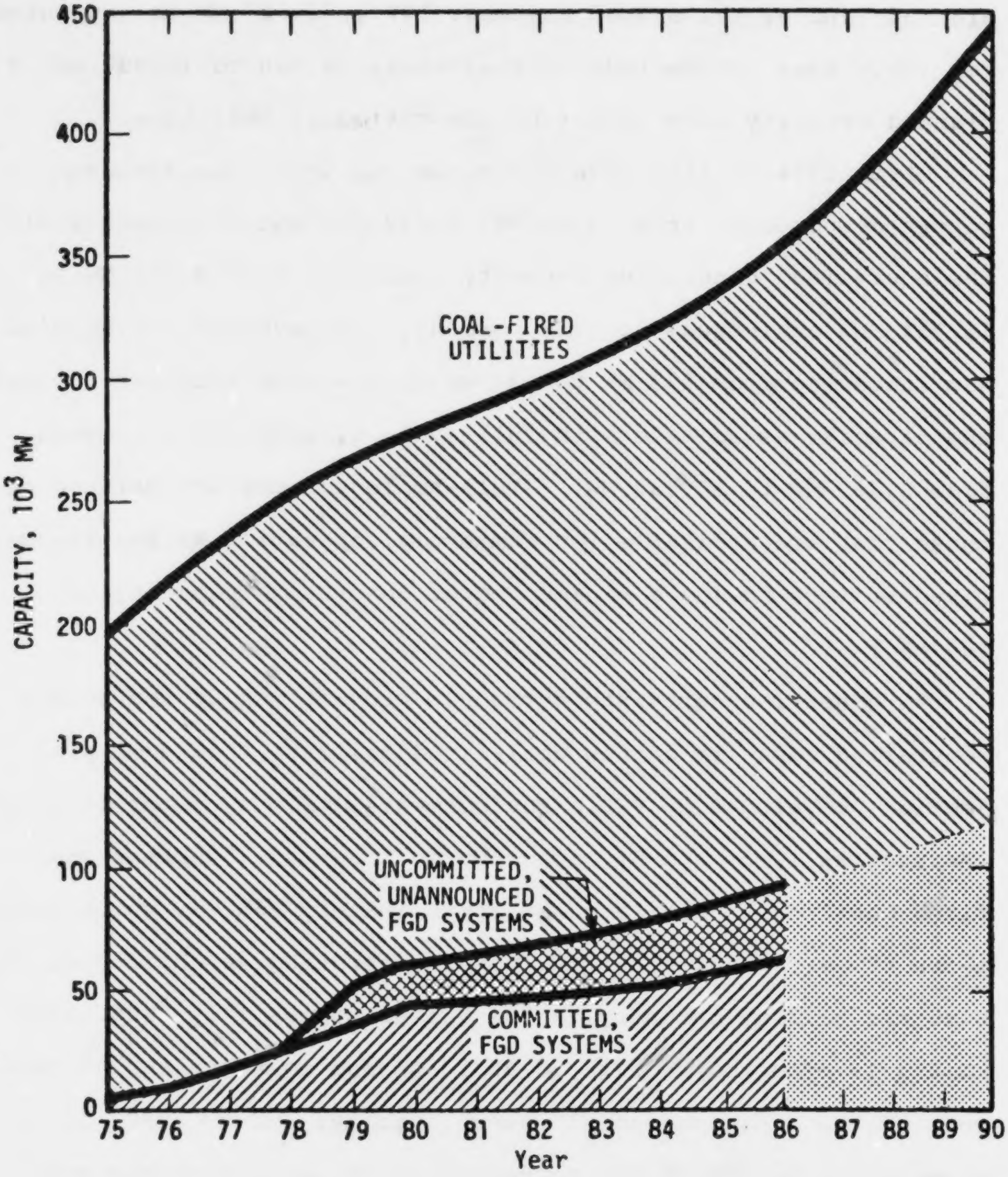

Figure 3. Projections of coal-fired generating capacity from 1975 to 1990 and FGD capacity from 1975 to 1986.

Table 1) are those that are currently in service, under construction, or planned. Virtually all of these systems are either currently operating or are committed to become operational at some future date. Uncommitted FGD systems are those that cannot be included in the committed group at this time because information regarding their status is not ready for public release. About 55 to 60 systems representing approximately 36,000 to 41,000 MW of generating capacity fall into this latter category. The premature stage of their planning, developments in on-going litigation, and the determination of applicable revised New Source Performance Standards (NSPS) and the Clean Air Act Amendments of 1977 generally preclude the inclusion of these systems. Although the revised sulfur dioxide NSPS had not been promulgated by the EPA at the time this paper was written, indications are that new coal-fired units that were not under construction by December 31, 1978 (requiring a construction permit by March 1, 1978) will be required to have some type of continuous control device. Impending NSPS covering sulfur dioxide are expected to be stringent enough to require the use of FGD technology because of its current level of commercial development. Thus, whereas approximately 6 percent of current total coal-fired power-generating capacity is now controlled by FGD, it is expected that about 25 percent of the total power-generating capacity from coal-fired facilities will be controlled by FGD by 1986.

NEW VERSUS RETROFIT

Figure 4 compares the application of new versus retrofit systems. Most of the initial applications were retrofitted. In 1975, for example, 60 percent of the operational FGD systems were retrofitted, whereas in 1980 about 70 percent will be on new systems. By 1985, about 75 percent of the operational FGD systems should be new unit applications. Any additional retrofits will stem primarily from systems required because of local regulatory action or small capacity demonstration projects.

PROCESS TYPES

The three primary methods are available for categorizing flue gas desulfurization processes: physical mechanism, chemical mechanism, and end-product mechanism. Physical mechanism refers to the phase in which sulfur dioxide removal is performed, i.e., wet or dry; chemical mechanism refers to the reagent used; and end-product mechanism refers to regenerable systems (in which sulfur dioxide is recovered in a usable, marketable form) and nonregenerable systems (in which sulfur dioxide must be disposed of as a nonrecoverable waste material). Table 2 summarizes, according to physical to mechanism, the systems that are operational, under construction, or for which a contract has been awarded.

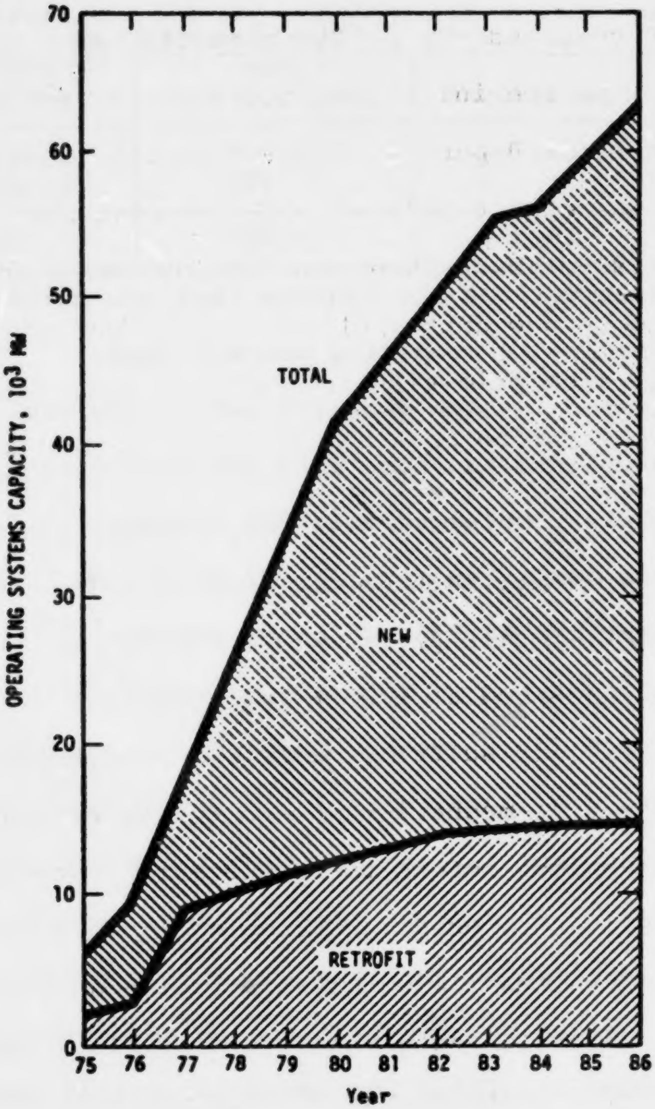

Figure 4. FGD operating capacity for new and retrofit installations through 1986.

TABLE 2. COMMITTED FGD CAPACITY BY PHYSICAL MECHANISM

Physical mechanism	FGD capacity, MW			
	Operational	Under construction	Contract awarded	Total
Wet	16,054	16,897	9,685	42,636
Dry	0	400	1,005	1,405
Total	16,054	17,297	10,690	44,041

Table 3 summarizes the systems that are either operating or planned, according to committed process type.

As indicated in both Tables 2 and 3, the vast majority of operating experience has been obtained with direct, calcium-based, wet-phase, nonregenerable FGD systems. Of the total active process-committed capacity of 50,377 MW, calcium-based systems account for 44,530 MW, or 88 percent.

Table 3 indicates an interesting trend in the utility industry's preference for limestone over lime processes. Limestone systems constitute approximately 59 percent of current calcium-based operating capacity, 59 percent of the capacity under construction, and 77 percent of that planned, or a total of about 65 percent. These figures show the industry's preference for limestone processes now and indicate an even stronger preference in those systems committed for operation within the next 5 years.

EMISSION LIMITING STANDARDS

Table 4 summarizes the FGD systems according to the regulatory standards they must meet. Of the 144 active systems, 65 (29,713 MW) are designed to meet the existing NSPS; 61 (26,567 MW) are designed to meet state standards more stringent than the

TABLE 3. DISTRIBUTION OF FGD SYSTEMS BY CHEMICAL PROCESS

Process	FGD capacity, MW			
	Operational	Under construction	Planned	Total
Limestone[a]	8,734	8,687	10,848	28,269
Lime[b,c]	6,070	6,029	3,482	15,581
Lime/Limestone	20	330	330	680
Sodium carbonate	375	509	0	884
Magnesium oxide	120	0	1,326	1,446
Wellman Lord	735	180	940	1,855
Dual alkali	0	1,102	0	1,102
Aqueous carbonate[d]	0	400	100	500
Citrate[e]	0	60	0	60
Total[f]	16,054	17,297	17,026	50,377

[a] Includes alkaline fly ash/limestone and limestone slurry process design configurations.

[b] Includes alkaline fly ash/lime and lime slurry process design configurations.

[c] Includes nonregenerable dry collection process design and nonregenerable wet scrubbing process design configurations.

[d] Includes nonregenerable dry collection process design and regenerable process design configurations.

[e] This system is being installed at St. Joseph Minerals' G.F. Wheaton Plant and is listed as a utility FGD system because the plant is connected by a 25-MW interchange to the Duquesne Light Company.

[f] Because the processes for all planned systems are not known, the totals in this table are less than those in Table 1.

existing NSPS* requirement; and 17 (5,427 MW) are designed to meet regulations less stringent than the existing NSPS. It is interesting to note that half of the 46 active, operational FGD systems (Table 1) are now meeting standards more stringent than the existing Federal NSPS.

TABLE 4. NUMBER AND CAPACITY OF ACTIVE FGD SYSTEMS FOR REGULATORY CLASSIFICATION CATEGORIES

Regulatory classification	Systems	Capacity, MW
Existing Federal NSPS	65	29,713
More stringent than existing Federal NSPS	61	26,567
Less stringent than existing Federal NSPS	17	5,427
Undetermined	1	800
Total	144	62,507

HIGH VERSUS LOW SULFUR COAL APPLICATION

The design and operation of FGD systems for high and low sulfur coal application represents another area of interest, especially with regard to the viability of such systems on boilers firing high sulfur coal. Because of the ambiguities inherent in the terms high and low sulfur coal, we have defined these as follows for purposes of this paper: low sulfur coal is any coal whose combustion will result in emissions equal to or less than 1.2 lb of sulfur dioxide per 10^6 Btu and high sulfur coal is any coal whose combustion will result in a higher emission value. Using these definitions, the following observations hold:

- Among the operating FGD systems, approximately 85 percent of the MW capacity is for high sulfur coal application.

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[*] The Clean Air Act NSPS value of 1.2 lb of sulfur dioxide per 10^6 Btu heat input to the boiler.

- Among the systems under construction, approximately 75 percent of the MW capacity is for high sulfur coal application.
- Among the planned systems, approximately 90 percent of the MW capacity is for high sulfur coal application.

SYSTEM SUPPLIERS

Approximately 30 companies offer FGD systems for application to utility boilers. Table 5 lists the companies that have supplied the systems that are in service, under construction, or under contract and identifies the number of systems and their capacities. Included in these totals are systems that are no longer used for sulfur dioxide removal.

INSTALLATION SCHEDULES

Schedules for the installation of FGD systems can vary greatly, primarily because of front-end activities relating to process selection and design. The period from startup to acceptance by the client can also vary depending upon system performance, and contractual agreements. The period between contract award and initial system start-up is less variable however. An analysis of the schedules of 35 systems indicates a range of 18 to 60 months with a mean of 32 months. The longer lead times are generally associated with new systems that are being constructed as an integral part of a new power generating system.

TABLE 5. MAJOR FGD SYSTEM SUPPLIERS

System supplier	Current status									
	Operational		Under Construction		Contract awarded		Terminated		Total	
	No.	MW	No.	MW	No.	MW	No.	MW	No.	MW
A.D. Little/Combustion Equipment Associates	6	1545	1	277	3	1900	1	20	11	3742
Air Correction Division, UOP	5	1540	2	934	1	720	2	220	10	3414
American Air Filter	3	667	3	985	0	0	0	0	6	1652
Babcock & Wilcox	2	1100	5	2069	4	1800	1	167	12	5136
Buell/Envirotech	0	0	1	575	0	0	0	0	1	575
Chemico	7	3745	2	800	1	527	2	245	12	5317
Chiyoda International[a]	1	23	0	0	0	0	1	23	2	46
Combustion Engineering[b]	9	3073	5	2675	3	1275	3	665	19	7703
Davy Powergas	3	735	1	180	2	940	0	0	6	1855
FMC	0	0	1	250	0	0	0	0	1	250
Mitsubishi International	0	0	0	0	2	980	0	0	2	980
Monsanto	0	0	0	0	0	0	1	110	1	110
Peabody	1	225	4	1625	0	0	1	163	6	2013
Pullman Kellogg[c]	0	0	3	1525	1	670	1	160	5	2355
Research Cottrell	5	2151	7	3203	1	1333	0	0	13	6147
Riley Stoker/Environeering	2	580	1	180	0	0	0	0	3	760
Rockwell International	0	0	0	0	1	100	0	0	1	100
United Engineers	1	120	0	0	2	1510	0	0	3	1630
Western/Niro	0	0	0	0	1	455	0	0	1	455
Wheelabrator-Frye/Rockwell	0	0	1	400	0	0	0	0	1	400

[a] The Scholz prototype was reactivated to demonstrate a new process design configuration. Thus, the entries in the operational and terminated categories refer to the same system.

[b] The original Lawrence 4 and 5 limestone injection systems were modified/replaced with a second-generation rod scrubber/spray tower design. Thus, the terminated category reflects the operational experience gained with these systems.

[c] The Mohave prototype is included in the terminated category although the supplier was not a participant in the Mohave Test Modules Program.

SECTION 3

GENERAL CONSIDERATIONS IN THE APPLICATION AND PERFORMANCE OF FGD TECHNOLOGY

Significant development of FGD technology in the United States dates from about the 1950's when bench-scale and limited pilot plant programs were initiated. Major pilot plant investigations first started in 1961, and between 1961 and 1978 more than 60 systems representing generating capacity of approximately 75 MW were investigated at the pilot plant level in the utility sector. Concurrent with later pilot plant investigations, prototype, demonstration, and full-scale systems were installed. The first commercial application of an FGD system on a utility boiler occurred in 1968. Since then, 62 systems representing a generating capacity of approximately 17,600 MW have been operated at the prototype, demonstration, and full-scale levels. As of the end of November 1978, 46 systems representing a generating capacity of 16,054 MW were in service, and another 98 systems representing 46,453 MW were under construction or planned.

Since the early investigations considerable progress has been in the development of FGD technology, and FGD is now considered to be the most commercially developed means of continuous control of sulfur dioxide emissions from coal-fired boilers.

The evolution of FGD technology from the limited pilot plant level to the full-scale commercial level can be attributed to several general process design and application considerations. Although it is difficult to quantify the impact that many of these factors have had on the development of the technology, an attempt is made to identify some of the general contributing factors in the balance of this section. (Section 4 provides more specific design and performance information on current technology trends.)

PROCESS DESIGN STRATEGY

Several general tendencies are evident in recent FGD design strategies. Generally, system designs incorporate an increased degree of flexibility and reliability. Specifically, trends are toward the sparing of modules and ancillary components and the designing of less interdependent systems (i.e., systems in which major unit operations are not strongly affected by upstream component performance).

SYSTEM APPLICATIONS

Many recent FGD facilities are installed on large base-loaded units designed to fire coal from a specific source. This generally results in a flue gas with more constant and stable characteristics, which can improve system reliability because the system does not have to respond to as dramatic a variation in flue gas flow rate and composition. In many of the original FGD applications, the systems were required to operate on widely varying loads (cycling and peak) and coal types (low sulfur western, high sulfur eastern, and blends); such situations often demanded response to conditions beyond their process control capability. As a result, variations in the reagent feed rate, loss of chemical control, and the incidence of chemical and mechanical problems caused numerous forced outages and lower dependabilities.

SYSTEM SUPPLIER AND ARCHITECTURAL-ENGINEERING EXPERIENCE

Later FGD system designs have benefitted from experience gained in the operation of first-generation systems. Building on this experience, system suppliers and designers are providing better process design configurations and materials of construction. That many suppliers now offer broader guarantees covering sulfur dioxide removal, particulate loading, mist loading, waste stream quality/quantity, power consumption, water consumption, reagent consumption, reheat energy consumption, and availability is indicative of this trend.

UTILITY EXPERIENCE

The utilities have also been gaining valuable operating and design experience. Many utilities have conducted or participated in FGD pilot plant programs and are thus better prepared to operate demonstration and full-scale systems. Operation of their first demonstration or full-scale system has also led to improved design and operation of subsequent systems.

REGULATORY AGENCY ATTITUDES

As FGD technology has evolved from a research, development, and demonstration effort to a means of continuous compliance with applicable regulations, local, state, and Federal regulatory agencies have changed their attitudes toward enforcement, compelling utility companies to improve the reliability of FGD systems.

PROCESS CHEMISTRY

Although scale and corrosion are still encountered and are sometimes still severe, general knowledge concerning scale formation and the occurrence of corrosion has greatly improved. As a result, systems are being designed and operated so that problems experienced by the earlier units will not be encountered.

SECTION 4

CURRENT TECHNOLOGICAL TRENDS

Considerable progress has been made in the development of conventional and emerging or advanced FGD processes. Much of this information has been acquired from the design and operation of first-generation FGD systems and translated into more effective designs and improved operation of newer systems. This section summarizes these emerging processes and presents a brief overview of their current status.

EMERGING PROCESSES

For the purposes of this discussion, processes within the emerging or advanced category are defined as those that incorporate major design and operating changes and thereby differ significantly from conventional direct lime/limestone processes. Of the processes so categorized, several have been evaluated at pilot and prototype development levels, and a few have progressed to the installation and operation of demonstration or full-scale units. Table 6 provides a brief summary of the emerging processes and highlights their current level of development and the extent of operating experience.

As is evident from Table 6, most of the previous operating experience has been with wet-phase sodium- and magnesium-based

TABLE 6. MAJOR EMERGING FGD [illegible] INVESTIGATED IN THE UNITED STATES

[illegible]	Developer	Current level of development	Previous operating experience	Remarks
ueous carbonate	Rockwell International	100-MW system (planned)	Mohave pilot plant test program	Full-scale application not yet demonstrated. 100-MW demonstration system scheduled for service in 1980.
Catalytic oxidation	Monsanto	100-MW system (terminated)	Wood River test program	No further process development.
Chiyoda Thoroughbred 101	Chiyoda International	20-MW system (terminated)	Scholz prototype test program	Development of the process has ceased in favor of a new design concept (Thoroughbred 121 which employs limestone reagent in a jet bubbler reactor).
Copper oxide adsorption	Shell/Universal Oil Products	Pilot plant	Big Bend test program	Process available for prototype or demonstration application. No systems planned at present.
Dual alkali	A.D. Little/ Combustion Equipment	277-MW system (construction)	Scholz prototype test program	Full-scale application not yet demonstrated. A 277-MW demonstration system scheduled for initial operation in early 1979.
	FMC	250-MW system (construction)	Industrial systems and utility pilot plants	Full-scale application not yet demonstrated. A 250-MW system is scheduled for service in 1979.
	Buell/Envirotech	575-MW system (construction)	Gadsby pilot plant test program	Full-scale application not yet demonstrated. A 575-MW system is scheduled for service in 1979.

TABLE 6. (Continued)

Process	Developer	Current level of development	Previous operating experience	Remarks
Magnesium oxide	Chemico	150-MW system (terminated) 95-MW system (terminated)	Mystic test program; Dickerson test program	Process demonstrated on full-scale oil- and coal-fired boilers. System now offered for commercial application.
	United Engineers	600-MW system (planned)	Eddystone test program	Eddystone 120-MW prototype test program still in progress. A full-scale 600-MW system is now planned for TVA's Johnsonville station.
Sodium carbonate	A.D. Little/ Combustion Equipment Associates	Three 115-MW systems (operational)	Reid Gardner Station	Three full-scale sodium carbonate (trona) FGD systems have been in service on coal-fired boilers at the Reid Gardner Station (Nevada Power). System performance has been good.
	Universal Oil Products	509-MW system (planned)	Jim Bridger pilot test program	A full-scale sodium carbonate (30% sodium carbonate purge solution from soda ash plant) FGD system is now being planned for the Jim Bridger station.
Wellman Lord	Davy Powergas	115-MW system (operational) 375-MW system (operational) 340-MW system (operational)	Crane test program	115-MW NIPSCO/EPA test program still in progress at Mitchell. Two full-scale systems have recently started operations at Public Service of New Mexico's San Juan Station.

TABLE 6. (Continued)

Process	Developer	Current level of development	Previous operating experience	Remarks
Dry adsorption	Foster Wheeler Bergbau Forschung	20-MW system (terminated)	Scholz prototype test program	No further development of system reported. Further evaluation of the sulfur reduction component (RESOX) in progress in Germany.
Dry collection	Wheelabrator-Frye Rockwell International	410-MW system (planned)	Leland Olds and Bowen Engineering pilot plant test programs	Successful testing has resulted in the planning of a full-scale, spray dryer and fabric filter FGD system at the Coyote Station.
	Joy/Niro	455-MW system (planned)	Leland Olds pilot plant test program	Successful pilot plant testing has resulted in the planning of a full-scale system at Antelope Valley. This system will use lime slurry as the reagent in a 2-stage atomizer/fabric filter design configuration.
	Babcock & Wilcox	550-MW system (planned)	Basin electric pilot plant test program	Successful pilot plant testing has resulted in the planning of a full-scale system at Laramie River. This system will use lime slurry as the reagent in a configuration which employs an ESP as the second stage collector.
	Carborundum/ Delaval	Pilot plant	Leland Olds pilot plant test program	Process similar in design to the Wheelabrator-Frye/R.I. and Joy/Niro processes. No full-scale applications have yet been announced.

processes that produce a recoverable, marketable byproduct or a high-quality filter cake. As might be expected, this substantial experience has resulted in the commercial application of dual alkali, Wellman-Lord, sodium carbonate, and magnesium oxide scrubbing systems. Three commercial dual alkali systems (Cane Run 6, A.B. Brown 1, and Newton 1) are now approaching startup. One demonstration (D.H. Mitchell 11) and two commercial (San Juan 1 and 2) Wellman-Lord systems are currently in service, and sodium carbonate systems have been operated commercially at three installations (Reid Gardner 1, 2, and 3). One demonstration magnesium oxide system (Eddystone 1) is currently in service, two such systems (Mystic 6 and Dickerson 3) have been terminated, and one full-scale system (Johnsonville) is planned.

The most recent promising development in emerging processes involves dry collection systems. Integrated processes designed to remove two or more pollutants simultaneously from the flue gas of a coal-fired boiler have commanded considerable interest for economic and operational reasons. The success of fabric filters in removing particulates from the flue gases of coal-fired boilers prompted investigations into the feasibility of using these filters to control both particulates and sulfur dioxide. Tests have been conducted using the injection of dry powdered nahcolite (a mineral form of sodium bicarbonate), calcium oxide, and calcium hydroxide into the flue gas stream and onto fabric filter bags for the removal of sulfur dioxide. Although impressive results were obtained with nahcolite, problems in meeting government requirements regarding nahcolite excavation in the producing areas of northern Colorado prompted research into other possible process design configurations. When it was found that using a spray dryer would eliminate the exclusive need for nahcolite, Wheelabrator-Frye and Rockwell International undertook a joint test program at Lelands Olds involving a two-stage system that combines a spray dryer and fabric filter. The spray dryer, the first stage, accomplished alkali injection and primary sulfur dioxide removal. The downstream fabric filter functioned as a second-stage sulfur-dioxide absorber and collection of flue gas particulates. Soda ash, trona, and lime, limestone, and fly ash slurries were tested as possible reagents in this system. Soda ash produced successful results. The data indicated that sulfur dioxide removals ranged from 48 to 98 percent at soda ash utilizations ranging from 96 to 65 percent for sulfur dioxide loadings ranging from 800 to 2800 ppm.

As a result of this successful testing, Wheelabrator-Frye and Rockwell International were awarded a turnkey contract for a full-scale system at Coyote 1, a 410-MW coal-fired unit.

Additional pilot plant testing involving different dry collection process design configurations has continued at Leland Olds, with systems supplied by Carborundum/DeLaval and Joy/Niro Atomizer. Babcock & Wilcox is also involved in a similar pilot plant evaluation of a dry-phase, two-stage collection system which is being conducted at another station. These programs involve the use of less expensive reagents, such as lime slurry, in various types of design configurations. Based on successful pilot plant testing, Joy/Niro has been awarded a contract for a full-scale dry collection system at Antelope Valley 1 of the Basin Electric Company and Babcock & Wilcox has been awarded a contract for a full-scale dry collection system at Laramie River 3 of the Basin Electric Company.

CONVENTIONAL PROCESSES

Conventional processes include all direct lime and limestone systems. Because these systems are the most widely applied, they are the ones with the most operating experience. Furthermore, they will have the greatest utilization in the very near future. For these reasons, lime and limestone systems have been subjected to extensive investigations, the results of which are summarized, along with general conclusions concerning process design.

Process Chemistry--Scaling, Sulfur Dioxide Removal, and Reagent Utilization

The use of reagents with low reactivity, such as limestone or lime, for sulfur dioxide removal from gas streams that can vary widely in flow and composition over short periods of time has resulted in several obvious and sometimes severe chemical limitations, including scale formation, low sulfur dioxide removal, and poor reagent utilization. A number of important findings regarding process chemistry and its effects on system performance have been made in recent years, and many of the process design and operating features that are now being incorporated into commercial systems are using these findings to improve performance. A brief summary of these features follows.

Two basic modes of lime/limestone FGD system operation will enable scale-free operation: coprecipitation and desupersaturation.

Coprecipitation involves the removal of calcium sulfate from the system as part of a calcium sulfite/sulfate solid solution. If a system is operated so that the maximum oxidation in the slurry circuit is about 16 percent, the scrubbing liquor remains subsaturated with respect to calcium sulfate (gypsum), and no hard scale occurs. If the degree of oxidation exceeds this level, more calcium sulfate is formed in the slurry circuit than can leave the system in a coprecipitated form. This causes the system to operate supersaturated with respect to gypsum. If relative saturation levels approaching the critical value of 1.4 is reached, the formation of hard scale can occur within the system.

Desupersaturation involves the removal of calcium sulfate from the system through the use of calcium sulfate (gypsum) seed crystals, which provide nucleation sites for the precipitation of calcium sulfate as well as through the calcium sulfite/sulfate solid solution. The seed crystals control sulfate scaling in a closed-loop system operating in a supersaturated mode. Crystal growth occurs on the seed crystals, the sulfate is removed from the system as gypsum, and relative saturation levels are kept below the critical 1.4 value.

Another approach to improving the chemistry of lime/limestone slurry systems so as to reduce scaling, increase sulfur dioxide removal, and improve reagent utilization is the use of additives in the slurry.

Use of Magnesium Additives--

Magnesium additives have proved to be effective. Increasing the magnesium ion concentration increases the liquid-phase alkalinity of the scrubbing slurry and increases the amount of sulfite and sulfate the scrubbing slurry can hold without exceeding solubility limits. The overall effect is a subsaturated operation that produces higher sulfur dioxide removal efficiencies and higher utilization. Experimental operating experience with magnesium additives has been accumulated at the Shawnee TVA/EPA Akali Scrubbing Test Facility and Paddys Run of Louisville Gas and Electric. Experimental and full-scale operating experience has been obtained at Phillips and Elrama of Duquesne Light, Bruce Mansfield of Pennsylvania Power, and Conesville of Columbus and Southern Ohio Electric.

Use of Other Additives--

The use of organic acids as additives is also under consideration. The use of carboxylic acids in lime/limestone scrubbing has been tested by the Tennessee Valley Authority (TVA) and TVA/EPA. This research has concentrated on the use of benzoic acid and, more recently, adipic acid at Shawnee. These acids are generally stronger than carbonic acid, but weaker than sulfurous acid. The addition of these acids has two effects: first, it aids mass transfer by buffering the pH of the liquid film at the gas/liquid interface; and second, because sulfurous acid is a stronger acid, the benzoate or adipate ion acts as a base in the sulfur dioxide absorption step. Thus, the addition of organic acid acids increases the total liquid phase alkalinity of the scrubbing liquor in much the same fashion as an increase in alkalinity because of magnesium ion. Intensive testing with adipic acid was recently performed at the TVA/EPA Shawnee alkali scrubbing test facility. In July 1978 initial test runs were performed without adipic acid to establish base lines for both lime and limestone scrubbing. These runs were followed by adipic acid testing, which continued throughout the balance of the year.

The preliminary results of this test program indicate agreement with initial expectations of higher sulfur dioxide removal efficiencies, and higher reagent utilizations. It has also been shown to be effective when used in conjunction with forced oxidation and when chlorides are present-conditions which adversely affect magnesium additives. One negative result, however, has been the unexpectedly high deterioration or decomposition of adipic acid that takes place in the scrubber. Actual feed rates of adipic acid were two to three times higher than could be accounted for in the system discharge sludge.

Design Changes--

A number of process design innovations also have been developed to eliminate scaling, increase sulfur dioxide removal efficiency, and improve reagent utilization. Forced oxidation is one technique that has been successfully piloted and used in commercial installations. On a number of systems the use of forced oxidation has contributed to scale-free operation, to meeting or exceeding design sulfur dioxide removals, and to achieving high reagent utilization levels and, most importantly, in improving the quality of the waste sludge. This has been especially true for limestone systems treating flue gas with low sulfur dioxide loadings. For these applications, the low sulfur dioxide levels coupled with the long liquid retention times result in high "natural" oxidation levels. This makes forced oxidation a particularly attractive method to improve sludge quality and minimize scaling, increase sulfur dioxide removal and improve reagent utilization in these systems.

Operating changes--

Several operating parameters have an important impact on scaling, sulfur dioxide, and reagent utilization.

Control of pH--Excluding all other factors, differences in optimum operating pH can affect the performance of lime/limestone slurry systems in two ways: operation at low pH generally promotes the formation of hard calcium sulfate scale (gypsum), and operation at high pH generally promotes the formation of softer calcium sulfite scale. Operating experience indicates that optimum pH levels are generally maintained between 8.0 to 8.5 for lime and 5.5 to 6.0 for limestone.

Solids level--If all other variables are held constant, increases in slurry solids levels will increase the amount of seed crystal area available for homogeneous crystallization. This is especially true for systems that control sulfate levels by desupersaturation. For these systems, an optimum amount of slurry seed crystals is maintained in the system by maintaining an optimum level of slurry solids. Thus, when the solids level drops, the seed crystal level drops correspondingly and causes the impairment or loss of homogeneous crystallization, the onset of heterogeneous crystallization, and subsequent scale development. Some of these systems aid desupersaturation by forcibly oxidizing all the sulfite to sulfate and then precipitate the calcium sulfate with the aid of gypsum seed crystals. Minor episodes of sulfate scaling have also occurred in these systems, in every case as a result of dilution of the slurry solids level after the mist eliminator wash water rate was increased to improve cleanliness. The increased levels of makeup water in the system decreased the slurry solids level and the seed crystal level, which impaired desupersaturation and resulted in scale formation. Reestablishment of slurry solids levels prevented further episodes of scaling.

Liquid-to-gas ratio (L/G)--If all other variables are held constant, increasing the L/G reduces the sulfur dioxide pickup per volume of scrubbing liquor. Thus, the relative saturation in the circulating slurry can be reduced by increasing L/G, assuming that desupersaturation takes place in the hold tank.

Process Chemistry--Corrosion

In simple terms, corrosion is the dissolving of metal surfaces. The incidence of corrosion in lime/limestone slurry FGD systems and design and operating measures taken to minimize or eliminate this problem are discussed briefly in the following paragraphs.

Two corrosive agents are present in the process: (1) sulfurous and sulfuric acids, and (2) chlorides. These two agents contribute to several specific types of corrosion: general corrosion, pitting, crevice corrosion, intergranular corrosion, stress-corrosion cracking, and erosion-corrosion.

A number of successful design, construction, fabrication, and operation measures have been developed to minimize the rate of corrosion or prevent it altogether. These measures are summarized briefly.

A selective process design approach developed by the major system suppliers allows highly corrosive environments to be isolated in discrete areas of the FGD system. This approach involves the separation of the scrubbing loop into separate multiple loops so that a different set of chemical conditions is maintained for quenching or prescrubbing, sulfur dioxide absorption, and mist eliminator washing (wash trays). In such designs, the quencher or prescrubber bears the first full brunt of the incoming hot flue gas. It encounters the total chloride content from the fuel fired without dilution and is the area in which low-pH, chloride-laden, return water is used freely in lime/limestone slurry systems. Isolation of such a corrosive environment to the quencher or prescrubber is advantageous because this area is small and discrete enough that it can be constructed of chloride-resistant materials without drastically increasing system cost. Alloys that have been tested and specified for full-scale systems are listed in ascending order of molybelmum content, pitting resistance, and cost: 317L stainless steel, Incoloy Alloy 825, Hastelloy G, Inconel Alloy 625, and Hastelloy C-276.

In many of the initial lime/limestone slurry systems, the incoming hot gas contacted the reactive absorbant suspension, which resulted in the accumulation of solids at the wet/dry interface. These deposits in some cases provided convenient sites for the accumulation of chloride at concentrations approaching 50,000 ppm. The result was severe episodes of pitting, stress corrosion, crevice corrosion, stress-corrosion cracking, and erosion-corrosion. This problem has been largely overcome by better control of process chemistry, use of self-cleaning devices, selective use of superior construction materials, and the use of multiple-loop designs. Better control of process chemistry eliminates the formation of scale in the system and thus prevents appearance of convenient chloride ion host sites. A number of systems are equipped with soot blowers in the approach ducts to the scrubber modules, which allows the inevitable buildup of solids at the wet/dry interface to be cleaned automatically and periodically.

Emission Control Strategy--

In recent years the trend in design of particulate and sulfur dioxide emission control systems has been toward combined electrostatic precipitator (ESP)/FGD or fabric filter (FF)/FGD strategies over simultaneous or two-stage wet scrubbing strategies. This preference is due to the high reliability afforded by ESP's and FF's, which enables selective bypass of scrubber modules without reduction of load or shutdown of the unit. Other benefits include the following:

- The potential for corrosion at wet/dry interfaces and erosion-corrosion in the FGD systems is minimized.
- Exotic construction materials can be used more selectively and in less amounts.
- Balanced-draft and booster fans can precede rather than follow the FGD system.
- Sludge blending and stabilization processes that use dry fly ash as an additive are premitted.

Equipment Design Improvements--

Specific design and operating improvements for FGD-related equipment are as follows:

Balanced-draft or booster fans--In addition to placement of these fans upstream of the FGD system, another development is the use of variable-pitch, axial flow fans. The main advantage of this design is its consistently higher efficiency (versus centrifugal fans) over the entire boiler operating range, which results in a substantial power savings. Other advantages are superior flow control, arrangement flexibility, easy access and maintenance, less severe construction requirements, and increased design reliability.

For scrubber modules, most suppliers now prefer 316 or 316L stainless steel because this material has demonstrated superior resistance to corrosion, errosion, and scale development compared with carbon steel, 304 stainless steel, and 304L stainless steel. This preference for 316 and 316L stainless steel is based primarily on the smooth mating surfaces and molybdenum content of these steels. The former attribute minimizes the presence of crevices that provide convenient sites for buildup of soluble chloride. The molybdenum content (2.50 to 2.75 percent minimum) of stainless steel increases corrosion resistance to localized attack such as pitting and crevice corrosion.

For mist eliminators, which have been very susceptible to corrosion, most suppliers now recommend the use of fiberglass-reinforced plastics, polypropylene, and corrugated plastics over stainless steels and other alloys because these materials are relatively lightweight, inexpensive, and do not corrode. Reheaters have been especially susceptible to corrosion, and the trend is toward indirect hot-air reheat because in-line reheat systems have been subject to corrosion and tube plugging.

Process Design

Several advances in the process design of lime/limestone slurry FGD systems have improved system dependability and sulfur dioxide removal. The following subsections describe these advances and some methods developed to reduce problems with major FGD equipment.

Dampers--Bypass and isolation dampers are used to regulate the flow of flue gas into and around the FGD system. The primary purpose of isolation is continuance of unit operation while the scrubber modules are under maintenance. Efficient and reliable dampers allow maintenance crews to service the modules in an efficient and timely manner. Common designs include slide-gate (guillotine), single-blade butterfly, and multiblade parallel (louver) dampers. Corrosion and erosion of the various types of dampers and damper seals have been common. In some cases, dampers have failed or been so inefficient that the modules could not be maintained during bypass situations. The current trend is toward two-stage louver dampers having a pressurized seal-air system that maintains a positive pressure between them. Pressurized seal air increases the energy demand of the system because of increased fan power requirements, but it contributes significantly to successful damper operation.

Scrubbers--Several recent design innovations have increased dependability and removal efficiency of scrubbers. Cooling the gas to its adiabatic saturation temperature prior to contact with the scrubbing slurry increases sulfur dioxide removal capability and minimizes the potential for scaling and corrosion at the slurry/gas interface area. Presaturators or quenchers were not incorporated into the design of many of the initial FGD systems. In such systems, the incoming, hot, pollutant-laden flue gas contacted the suspended reactive absorbent and resulted in solids accumulation and subsequent corrosion. For this reason, presaturators or quenchers that use clear liquor or spent slurry for the absorbing stage (multiple slurry loops) are now used (or plans call for their use) in systems that include dry-phase particulate precollection.

The trend in the design of lime/limestone slurry FGD systems is away from venturis and packed beds to spray towers and combination towers. The venturi design was abandoned largely because the small liquid/gas contact time results in relatively low sulfur dioxide absorption. Scaling, plugging, and corrosion of internals occurred in some of the packed-bed designs (fixed and mobile) and tray towers. Spray towers, on the other hand, have few internal components in the gas/liquid contact zone and therefore offer the potential for greater dependability because there are fewer sites for deposition of solids in the form of scale, collected fly ash, and unused reagent. To date, spray tower operation has been very successful. Although high dependability and sulfur dioxide removal have been reported for almost all the FGD systems incorporating spray tower designs, several limitations also have been encountered. Mass transfer limitations tend to restrict spray tower design applications so that only low- and medium-sulfur coal can be used in conventional lime/ limestone slurry systems.* In addition, the greater tendency for slurry carryover in spray towers requires either increased tower height or special mist eliminator designs (wash trays, bulk entrainment separators), both of which increase capital and annual cost requirements.

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*When high-sulfur coal is burned, spray towers in service and scheduled for operation use or plan to use special reagents (carbide lime, magnesium-promoted lime, and limestone) to compensate chemically for mass transfer limitations.

These limitations have given rise to the development of combination towers. These towers combine the features of venturi, packed, tray, and spray towers into one module. Examples of combination tower designs now offered by some of the major system suppliers include spray/packed towers, venturi/spray towers, and tray/packed towers. These designs offer greater flexibility because extreme operating conditions can be segregated into discrete areas of the scrubber, allowing separate chemical and physical conditions to be maintained. This permits the use of the two-loop slurry concept, in which low pH liquor contacts the entering flue gas in an initial scrubbing loop, where some sulfur dioxide removal takes place. High pH liquor is contacted with the gas in the second scrubbing loop, where the bulk of the sulfur dioxide removal takes place. Spent slurry from this loop is discharged to the first loop where the unused reagent is consumed. Fresh makeup reagent is added only in the second loop. This type of design takes advantage of the concept of contacting the flue gas containing the highest sulfur dioxide concentration with the lowest liquor alkalinity and the highest liquor alkalinity with the lowest sulfur dioxide concentration. Performance has verified the potentially high removals and utilizations afforded by such designs.

Reaction tanks--Coinciding with gas-side staging is liquid-side staging, in which hold tanks are arranged in series to simulate plug flow reactor designs. (A plug flow design is one that allows the reacting liquor to flow through the reactor without backmixing. A plug flow situation can be approximated by arranging agitated tanks in series.) This design concept was originally piloted at IERL-RTP and further tested at Shawnee. A number of full-scale systems that incorporate liquid-side staging have resulted; all are low-sulfur, limestone-slurry systems. Sulfur dioxide removals and dependabilities greater than 90 percent have been reported.

Mist elimination--Chevron and baffle-type mist eliminators continue to be the only designs used in U.S. utility FGD systems. Several different designs have been tested (including wire-mesh, tube-tank, gull-wing, ESP, and radial vane), but the performance and economics associated with these and other design alternatives indicate that the exclusive use of chevron and baffle types will probably prevail. The popularity of these separators is due primarily to design simplicity and flexibility, adequate collection efficiency for medium to large size drops, relatively low pressure drop, open construction, easy access for maintenance, and relatively low cost.

Within these two preferred types of mist eliminators, a number of specific design, construction, and operation improvements have been implemented, with the following results:

1. Chevron designs of continuous-vane construction now predominate over noncontinuous-vane construction because of their greater strength and lower cost.

2. Multiple-stage designs predominate over single-stage designs on limestone systems. This tendency is process-sensitive in that limestone systems, which outnumber lime systems in the United States, generally require two or three stages for effective mist entrainment separation.

3. Single-stage designs are successful for lime systems because of the superior reactivity of lime and the correspondingly higher utilization.

4. The number of passes per stage also tends to be process-sensitive. Four-pass designs are generally used for lime and three-pass designs for limestone. More passes are required for lime systems because the single-stage design is used, whereas fewer passes are required for limestone systems because the multiple-stage design is used.

5. Fiberglass-reinforced plastics, polypropylene, and corrugated plastics are now used in almost every operational system and specified for use in nearly every planned system. These materials are preferred because they are relatively lightweight, inexpensive, and superior in resistance in corrosion. Potential problems associated with high temperature excursions have been minimized by specifying materials that can withstand exposure up to 400°F.

6. Vane spacings of 1.5 to 3.0 in. are generally used in single or first stages and 0.9 to 1.0 in. for second stages. Multiple staging permits the use of finer spacings, which provides increased mist-separation capability for smaller particles.

7. The horizontal configuration (vertical gas flow) is still widely used because of its adequate performance (to date), its operational and design simplicity, and its lower capital cost. The vertical configuration offers a number of advantages over the horizontal configuration. For example, reentrainment due to the gas flow opposing the path of the drainage is eliminated, and limitations on wash water quality and quantity (as well as wash direction) are eliminated. Two systems recently started operations using vertical configurations. Initial results indicate adequate operation performance and no major operating problems.

8. Special features, such as hooks and pockets on the vanes, are desirable for prevention of reentrainment.

9. Bulk separation devices, impingement plates, single baffle deflectors, and gas direction changes are becoming integral parts of mist eliminators because they increase removal efficiency and design flexibility.

10. Wash and knock-out trays have been incorporated into a number of mist eliminators to conserve freshwater and increase (or extend) the quantity of water available for washing.

11. Wash systems that use blended water consisting of pond return water or thickener overflow and freshwater are used over other strategies (total return or total makeup). Intermittent, high-pressure, high-velocity wash systems are preferred to continuous wash systems because they have less impact on water balance and chemistry.

12. Optimum distances between stages are generally 4 to 5 ft, and freeboard distances are 4 to 5 ft. The former is the minimum distance permitting easy access for maintenance. The latter is the distance at which carryover can be minimized without drastically increasing tower height and pressure drop.

13. Superior overall operation is obtained when fly ash is collected prior to the scrubbing system. This is because scrubbing systems in which fly ash is not used usually have a low slurry solids content. The lower the slurry solids content, the less likely the tendency for mist eliminator fouling.

Reheaters--A pronounced preference for stack gas reheat versus no reheat (wet stack) is still evident for those systems in service and committed for future operation. The use of wet stacks at a number of systems has been abandoned in favor of reheat because problems encountered with corrosion, plume dispersion, and plume visibility. Such problems are more pronounced where high-sulfur coal is burned because sulfur dioxide loadings are higher. A number of developements concerning reheater design and construction are noted:

1. Six methods are available to increase the temperature of the gas from a scrubber prior to discharge to the stack: in-line reheat, direct combustion reheat, indirect hot air reheat, gas bypass reheat, exit gas recirculation reheat, and waste heat recovery reheat.

2. Of the six methods now available only four are applied in commercial operations in the United States: in-line reheat, direct combustion reheat, indirect hot air reheat, and gas bypass reheat.

3. Among the systems that have operated or are currently in service, in-line reheat has proved to be the most popular strategy.

4. The trend in reheat strategies, as evidenced by FGD systems scheduled for immediate and future operation, is away from in-line and direct combustion methods and toward indirect hot air reheat. This is largely due to the problems encountered with in-line reheaters and the need for oil or natural gas with direct combustion reheaters. In-line reheat systems have been subject to corrosion and plugging in the tubes. The corrosion in many cases has been so severe that even the heartier alloys have been unsatisfactory under many operating conditions. Many of these problems have been attributed to upstream mist eliminator inefficiency and inadequate self-cleaning techniques (soot blowers).

5. A number of the major system suppliers still recommend in-line reheaters, especially when minimization of the energy demand is desired. It has been determined that corrosion of high-alloy materials is attributed to stress corrosion caused by chloride, whereas carbon steel is more susceptible to acid corrosion caused by sulfur dioxide. Therefore, if low sulfur/low chloride, low sulfur/high chloride, or high sulfur/low chloride environments can be accurately predicted, in such applications in-line reheaters may be used successfully.

6. Indirect hot air reheat has the undesirable effect of increasing the energy demand of the FGD system and increasing overall system cost.

7. Bypass reheat may be used for low-sulfur coal FGD applications when the required degree of reheat is not seriously constrained by emission standards.

8. The use of efficient mist eliminators reduces the load on the reheat system by removing water droplets from the flue gas stream.

9. ΔT's of 14° to 28°C (25° to 50°F) adequately prevent downstream water condensation.

10. Waste heat recovery reheat (regenerative) is now being specified for two full-scale systems planned for future operation. In this system, the sensible heat of the incoming flue gas is recovered in an in-line heat exchanger placed upstream of the air preheater. This heat is then used to reheat the scrubbed gas stream. The in-line heat exchanger can be a direct gas-gas heat exchanger or a gas-liquid heat exchanger that uses a fluid of high heat capacity. Experience with these systems has been reported for experimental, small-scale, pilot plant (1 MW) tests.

Solids separation (sludge dewatering)--The major development in this area is the increased emphasis placed on clarification, centrifugation, and vacuum filtration, and the corresponding decreased emphasis on interim ponding. Formerly, an interim pond was relied on to fulfill three functions: clarification, dewatering, and temporary or final sludge storage. The realization that a single pond cannot perform all three functions adequately encouraged the development of the other techniques. Furthermore, the increased emphasis placed on offsite disposal for landfill and structural fills and on attaining closed water-loop operations also stimulated the use of clarifiers, centrifuges, and vacuum filters. In addition to using these techniques, several installations are using forced-oxidation strategies to enhance solids settling and filtration properties, to improve process chemistry, to improve waste sludge quality, and to decrease land requirements for sludge disposal.

Process Control and Instrumentation--

Because of the complex nature of lime/limestone scrubbing chemistry, which has been the primary source of operating problems in full-scale systems, process control is considered a crucial item. The following is a brief review of some of the essential findings and innovations in the development of process control technology:

1. Virtually all of the operating full-scale systems regulate reagent feed rate by controlling slurry pH. A pH sensor provides a signal for modulating the flow of reagent to the FGD system in a feedback control mode. The pH signal regulates the position of control valves for controlling the rate of reagent feed.

2. The major problems encountered with pH control systems are sensor plugging, calibration drift, breakage, false indication, and erosion/corrosion damage.

3. Sufficient operating experience has been obtained so that most of the reagent feed control problems have been identified. Once identified, these problems have been resolved for the most part through design modifications and/or new operating and maintenance procedures.

4. Concerning selection of hardware for pH control, it has been noted that dip-type sensors are more successful than in-line sensors because they are easier to clean and calibrate. In-line, flow-through sensors are generally subject to more wear and abrasion and generally require more frequent maintenance.

5. Other reagent feed control systems have been or are being evaluated on full-scale systems. One type involves feed-forward reagent control on the inlet flue gas flow rate and sulfur dioxide concentration, with trim provided by slurry pH. Another type involves control of reagent feed rate by using the outlet sulfur dioxide as the control variable. Limited success has been reported on both of these systems, primarily because of the difficulty in obtaining accurate and consistent readings from sulfur dioxide gas analyzers. This has been especially difficult on high-sulfur coal applications.

Construction Materials--

Analysis of FGD experience in the United States to date, when examined at all levels of development, makes one point clear: generalization is difficult because of the many factors and apparent contradictions in FGD operation. This is especially evident in the area of construction materials. Many examples can be cited in which seemingly inferior construction materials have been adequate, whereas apparently adequate materials have failed. Although construction materials have been discussed throughout this paper with respect to process and equipment design improvements, a brief review of the trend in the construction of critical elements in FGD systems is provided as follows:

1. The use of 316 and 316L stainless steel is generally preferred as the construction material in critical areas of the FGD system.

2. Some designers avoid 316 and 316L stainless steels whenever possible and instead use carbon steel with a surface lining or coating that physically shields the bare metal from the corrosive environment. These linings are usually resins applied in liquid or semiliquid form, by spray or trowel, and allowed to cure. The use of lined or coated carbon steel offers the potential benefits of being able to withstand low pH/high chloride environments much better than the 316's and being considerably less expensive. Actual operating experience, however, has shown that these materials are very susceptible to high temperature excursions, and often require an additional capital investment for an auxiliary power system. Corrosion resistance is also a limiting factor for many of the materials used. In addition, application and reapplication of linings have been suspect, especially reapplications, where proper preparation of the metal surface becomes more difficult.

3. The use of chemically resistant masonry materials (e.g., silicone carbide) and ceramic liners in the throat areas of venturi scrubbers (usually applied to 316 or 316L stainless steel at the converging section of the venturi, where the gas velocity and erosive nature of the fly ash are highest), has been quite successful.

4. The use of natural rubber, neoprene, polyvinyl chlorid (PVC), fiber-reinforced plastic (FRP), and flaked-glass polyester generally predominates in the liquid and thickener circuits of the FGD system (tanks, pumps, agitators, piping, and thickeners) where the metal parts are 316 stainless steel or Alloy 20. For filtration systems, neoprene, polypropylene, FRP, and Alloy 20 are generally employed.

5. Many systems are also being constructed of more resistant alloys in trouble spots (e.g., wet/dry, high-temperature, and high-chloride environments such as a prescrubber or presaturator). Alloys such as Hastelloy C-276, Hastelloy G, Alloy 20, Inconel 625, Incoloy 825, 317 low-carbon stainless steel, 904 low-carbon stainless steel, Jessop JS-700, and E-Brite 26-1 are being selected in minimum amounts.

6. The use of stack lining or coating materials has been troublesome, especially where high-sulfur eastern coal is burned. Lining failures have also been reported in systems that included an apparently adequate degree of reheat. Recent developments in this area indicate that some progress has been made. A proprietary, spray-on, elastomer has been applied on three of the four flues at a station burning high-sulfur eastern coal (without reheat) with apparent success. The use of acid brick also appears to be successful in a similar application. At installations burning low-sulfur coal, adequate linings have not been as great a problem; many installations can get by with unprotected carbon steel flues.

SECTION 5

CAPITAL AND ANNUAL COSTS OF OPERATIONAL FGD SYSTEMS

INTRODUCTION

The cost of FGD systems is an area of intense interest and substantial controversy. As a result, a number of computer models have been developed in recent years to estimate capital and annual costs. In an effort to provide meaningful economic data on FGD systems, PEDCo Environmental has incorporated reported economic data into the EPA Utility FGD Survey Report. This information has appeared as a separate appendix to these reports since October 1976. Until May 1978, this cost appendix consisted entirely of data reported by the utilities, and little or no interpretation was provided by PEDCo Environmental. Beginning with the May 1978 report, however, the format and content of the cost appendix were revised to include adjusted costs for the operational FGD systems.

APPROACH

In March 1978, a cost form was forwarded to each utility having an operational FGD system. The form contained all the available cost information for the system or systems at that particular utility. Having been notified in advance of the purpose of the project, each utility was requested to review the cost form thoroughly. A followup visit by the PEDCo Environmental staff was arranged to assist in data acquisition and to ensure completeness and reliability of information. Results of the cost analysis were then forwarded to each participating utility for final review and comment.

Analysis of the cost data centered on adjusting the estimates to a common basis. The data were analyzed solely to determine accurate costs of FGD systems, not to critique the design or reasonableness of the costs reported by any utility. The primary adjustments were as follows:

- All capital costs were adjusted to July 1, 1977, dollars using the Chemical Engineering Index. All capital costs, represented in $/kW, were expressed in terms of gross megawatts (MW).
- Particulate control costs were deducted. Since the purpose of the study was to estimate the incremental cost of sulfur dioxide control, particulate control costs were deducted, using either data contained in the cost breakdowns or a percentage of the total direct cost (capital and annual). The percentage reduction varied according to system design and operation.
- The capital costs associated with the modification or installation of equipment not directly involved with the FGD system were included (e.g., stack lining, modification to existing ductwork or fans).
- Indirect charges were adjusted, where necessary, to provide adequate funds for engineering, field expenses, overhead, interest during construction, startup, and contingency.
- All annual costs, given in mills/kWh, were based on net MW.
- All annual costs were adjusted to a common capacity factor (65 percent).

- Replacement power costs were not included because only a few uitlities reported such costs and those reported were presented in a variety of methods.

- Sludge disposal costs were adjusted to reflect the costs of sulfur dioxide scrubber sludge disposal only (i.e., excluding fly ash disposal) and to provide for disposal over the anticipated lifetime of the FGD system.

- A 30-year life was assumed for all process and economic considerations for all new units. A 20-year life was assumed for all process and economic considerations for all retrofit systems, even if the remaining boiler life was less than 20 years.

Cost data were obtained for 27 of the 31 FGD systems in operation at the time this cost analysis was conducted. Table 7 provides a summary of the reported and adjusted capital and annual costs for all the operational FGD systems, and Table 8 summarizes these results by category (process) and application (new/retrofit).

TABLE 7. REPORTED AND ADJUSTED CAPITAL AND ANNUAL COSTS FOR OPERATIONAL FGD SYSTEMS

	Reported		Adjusted	
	Capital, $/kW	Annual, mills/kWh	Capital, $/kW	Annual, mills/kWh
Cholla 1	52.0	2.19	56.0	2.58
Conesville 5	55.6	4.71	70.8	7.42
Elrama 1-4	113.5	8.62	127.2	7.81
Phillips 1-6	107.0	7.83	140.6	8.57
Petersburg 3[a]	99.5		100.6	6.56
Hawthorn 3-4	29.3	8.40	87.3	4.35
La Cygne 1	53.7	1.70	68.0	3.78
Green River 1-3	70.3	14.35	77.6	5.24
Cane Run 4	66.6	2.75	80.6	5.78
Cane Run 5[a]	62.4		67.5	5.56
Paddys Run 6[b]	52.9		76.5	6.51
M.R. Young 2[a]	86.0		93.1	5.16
Colstrip 1-2	77.1	0.27	77.3	4.06
Reid Gardner 1-2	42.9	2.10	60.9	3.20
Reid Gardner 3	113.6	2.10	107.9	4.38
D.H. Mitchell 11	156.9	14.86	145.5	12.73
Sherburne 1-2	47.9	1.99	71.9	2.77
B. Mansfield 1-2	120.7	14.35	102.9	8.68
Eddystone 1A[c]	156.8		233.3	
Winyah 2	47.5	1.61	66.5	2.92
Southwest 1[a]	77.3		117.7	6.17
Widows Creek 8	98.2	2.99	113.2	5.28

[a] Annual costs were not reported by the utility for this system because it has not had operating status long enough to provide.

[b] Annual costs were not reported by the utility for this system because the peak load status of unit precludes its providing meaningful data.

[c] Annual cost data are being assembled by the utility.

TABLE 8. CATEGORICAL RESULTS OF THE REPORTED AND ADJUSTED CAPITAL AND ANNUAL COSTS FOR OPERATIONAL FGD SYSTEMS

	Reported						Adjusted					
	Range, $/kW	Capital average, $/kW	σ	Range, mills/kWh	Annual average, mills/kWh	σ	Range, $/kW	Capital average, $/kW	σ	Range, mills/kWh	Annual average, mills/kWh	σ
All	29.3-156.9	78.0	35.7	0.27-14.86	5.6	5.1	56.0-233.3	94.2	36.9	2.58-12.73	5.5	2.4
New	47.5-120.7	78.8	27.8	0.27-14.35	4.3	5.4	66.5-117.7	86.8	17.8	2.77-8.68	5.2	2.1
Retrofit	29.3-156.9	77.2	42.9	2.10-14.86	7.4	5.4	56.0-233.3	101.0	48.2	2.58-12.73	5.8	2.7
Nonregenerable	29.3-120.7	71.7	28.7	0.27-14.35	5.2	4.8	56.0-140.6	86.5	22.3	2.58-8.68	5.2	1.9
Regenerable	156.8-156.9	156.8	0.1	14.86	14.9		145.5-233.3	189.4	62.1	12.73	12.73	
Limestone	47.5-99.5	71.4	23.7	1.61-2.99	2.1	0.6	56.0-117.7	87.0	26.7	2.58-6.56	4.5	1.7
Lime	29.3-120.7	75.3	34.6	2.75-14.35	9.3	4.3	67.5-140.6	92.8	23.5	4.35-8.68	6.6	1.7
Alkaline fly ash/limestone	47.9	47.9		1.99			71.5	71.5		2.77	2.77	
Alkaline fly ash/lime	77.1-86.0	81.6	6.3	0.27	0.27		77.3-93.1	85.2	11.2	4.06-5.16	4.6	0.8
Sodium carbonate	42.9-113.6	78.3	50.0	2.10	2.1		60.9-107.9	84.4	33.2	3.20-4.38	3.8	0.8
Magnesium oxide	156.8	156.8					233.3	233.3				
Wellman-Lord	156.9	156.9		14.86	14.86		145.5	145.5		12.73	12.73	

SECTION 6

SUMMARY

The preceding sections of this report examined the entire gamut of FGD technology in the utility sector of the United States. As discussed in Sections 3 and 4, considerable progress has been made in resolving the major problems that have plagued the initial installations. More work needs to be done, however, especially in the area of evaluating trade-offs as a means of reducing cost and parasitic energy demand (power production penalty) without impairing dependability and efficiency. The report implies a number of research, development, and demonstration (RD&D) needs that should be pursued to increase the efficiency and cost-effectiveness of FGD operation. The main areas for RD&D work are as follows:

- A thorough investigation of the phenomenon of scaling. The mechanisms that are involved in scaling and the border zone between scaling and nonscaling should be better defined.
- Improvement of mist elimination. Because this is a particularly troublesome area, an intensive investigation aimed at optimizing design would be beneficial.
- Optimization of stack gas reheat. A decision regarding whether or not reheat is required, and if so, the methods that provide the best results (energy, equipment protection, plume visibility, and mechanical reliability) would be beneficial.

- Improvement of instrumentation and process control strategies. This would be very beneficial to cost-effective and reliable operation.

- Optimization of construction materials used in the FGD system and related equipment. Recent disastrous results with alloys and linings (especially the latter) used on the stack accentuate the need for a major effort in this area.

- Closed-water loop operation. This needs work as far as ascertaining overall feasibility and determining the best applicable methods.

- Secondary environmental impacts, especially from sludge resulting from nonregenerable systems, need further investigation. More work is needed to reduce land requirements and increase utilization of the material in land and structural fills.

- Process complexity and economic and energy penalties associated with the current generation of regenerable systems. It would be beneficial if these could be minimized.

- Investigation of dry collection processes. A substantial amount of investigation has already been conducted to verify process design using lime and sodium carbonate reagents for low- to medium-sulfur coal applications. The feasibility of dry collection for medium- to high-sulfur coal applications could prove beneficial and should be investigated.

- Investigation of process design configurations in which the ESP follows rather than precedes the scrubber. Such configuration may offer advantages for particulate collection. This may prove beneficial in light of recent concerns over scrubber slurry carryover and mist loading in the scrubbed gas stream and compliance with more stringent particulate emission regulations.

REFERENCES

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2. Temple, Barker, and Sloan, Inc. Policy Testing Model for Electric Utilities, Exhibit II-3.

3. Twelfth Annual Power Engineering Survey. Power Engineering, April 1978.

RECENT RESULTS FROM EPA'S LIME/LIMESTONE SCRUBBING PROGRAMS -ADIPIC ACID AS A SCRUBBER ADDITIVE-

by

H.N. Head, S.C. Wang, and D.T. Rabb
Bechtel National, Inc., 50 Beale Street
San Francisco, California 94105

and

R.H. Borgwardt, J.E. Williams, and M.A. Maxwell
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

ABSTRACT

Adipic acid has been demonstrated as a powerful scrubbing additive for enhancing SO_2 removal in lime and limestone wet scrubbing tests both at EPA's IERL pilot plant at Research Triangle Park, North Carolina, and at the EPA-sponsored Shawnee Test Facility near Paducah, Kentucky.

At adipic acid concentrations in the scrubber liquor in the range of 700 to 1500 ppm, SO_2 removals in excess of 90 percent have been consistently obtained with limestone slurry at both facilities. SO_2 removal was effectively enhanced even when operating with dissolved chlorides in the scrubber liquor as high as 10,000 ppm at the Shawnee Test Facility and 17,000 ppm at the IERL-RTP pilot plant. SO_2 removal was enhanced equally well in systems with or without forced oxidation. Adipic acid was found to cause only minor differences in the dewatering and handling properties of oxidized sludge. No scaling problems were encountered.

Because of decomposition at scrubber operating conditions, consumption of adipic acid has been greater than anticipated, especially in systems with forced oxidation. To maintain 1500 ppm of adipic acid in the slurry liquor at Shawnee conditions, adipic acid consumption has been in the range of 8 to 9 lb/ton of limestone.

A preliminary economic assessment by TVA comparing a base limestone case with adipic acid enhanced limestone for 90 percent SO_2 removal or better and using the actual adipic acid consumptions experienced at Shawnee indicates a reduction in annual revenue requirements of about 0.3 to 0.4 mill/kWh exclusive of ponding costs for systems with 750 to 1500 ppm of adipic acid.

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NOTE

Although it is the policy of the EPA to use the metric system for quantitative descriptions, the British system is used in this report. Readers who are more accustomed to metric units are referred to the conversion table in the Appendix.

RECENT RESULTS FROM EPA'S LIME/LIMESTONE SCRUBBING PROGRAMS -ADIPIC ACID AS A SCRUBBER ADDITIVE-

Section 1

INTRODUCTION

A primary objective of the EPA alkali wet scrubbing test program during the last year has been to enhance SO_2 removal and improve the reliability and economics of lime and limestone wet scrubbing systems by use of adipic acid as a chemical additive. This report addresses the results of testing with adipic acid at EPA's IERL-RTP pilot plant and at the EPA Alkali Scrubbing Test Facility, near Paducah, Kentucky.

Adipic acid is a commercially available organic acid, [$HOOC(CH_2)_4COOH$], that buffers the pH* in SO_2 absorbers when present at low concentrations in the scrubbing liquor. The theoretical basis for its effect on the performance of limestone and lime scrubbers was developed in detail by Rochelle1): The buffering action limits the drop in pH that normally occurs at the gas/liquid interface during SO_2 absorption, and the higher concentration of SO_2 in the surface film resulting from this buffering action accelerates the liquid-phase mass transfer. The capacity of the bulk liquor for reaction with SO_2 is also increased by the presence of calcium adipate in solution. SO_2 absorption is therefore less dependent upon the dissolution of limestone in the absorber. The overall result of these effects is improved SO_2 removal in limestone or lime scrubbers of a given type operating at a given L/G. In the case of limestone, it follows that a given SO_2 removal efficiency can be achieved at a lower stoichiometric ratio.

Further analysis by Rochelle2) indicated that the use of additives would be most attractive economically when used in scrubbers employing forced oxidation. If no decomposition or volatilization of the additive occurs, the makeup requirements for the additive would be minimized by the tightly closed loop resulting from the better dewatering properties of oxidized sludge. For this reason, the scrubber tests reported here emphasized the evaluation of adipic acid in combination with forced oxidation.

Adipic acid has several potential advantages over other additives, such as MgO, which are also known to increase SO_2 removal, but do so by means of reactions involving the sulfite/bisulfite equilibrium. Since adipic acid does not depend on this mechanism for its buffering activity, it is not adversely affected by oxidation of the sulfite in the scrubbing liquor. It should therefore be especially useful in single loop scrubbers that employ forced oxidation. Furthermore, the buffering mechanism by which adipic acid enhances SO_2 absorption is not affected by the presence of chloride. The lack of interference by chloride means that adipic acid should be fully effective in the most tightly closed loop systems.

---

* Adipic acid has two buffering points. In the absence of chloride these are at pH 5.5 and 4.5. Chloride concentrations in the range of 5000 to 7000 ppm depress these buffering points to about pH 5.0 and 4.0.

Preliminary economic evaluations have shown that adipic acid can reduce the operating cost of limestone systems while simultaneously increasing performance. Adipic acid is a major ingredient in the manufacture of nylon and is sufficiently available that widespread use in commercial FGD systems would have little effect on the market.

Beginning in 1977, initial studies conducted by EPA with the 0.1 MW pilot plant at the Industrial Environmental Research Laboratory located at Research Triangle Park, North Carolina (IERL-RTP) demonstrated as predicted by Rochelle that adipic acid was indeed an attractive additive. Results of these tests were first reported at the EPA Industrial Briefing at Research Triangle Park in August 1978.(3) An update of recent findings at the pilot plant is given in Section 2.

Based on the findings at the IERL-RTP pilot plant, a program was set up at the EPA sponsored Test Facility located at the TVA Shawnee Steam Plant near Paducah, Kentucky, to develop commercially usable design data for adipic acid as a chemical additive. Testing with adipic acid was initiated in July 1978 on the 10 MW EPA prototype scrubbers and has continued since as a major part of the Shawnee Advanced Test Program. The results of testing with adipic acid additive at the Shawnee Test Facility from July 1978 through January 1979 are presented in Sections 3 and 4.

Configurations successfully demonstrated during this period were:

- Adipic acid enhanced limestone scrubbing in a venturi/spray tower system with two scrubbing loops and forced oxidation in the first loop.
- Adipic acid enhanced limestone scrubbing in a Turbulent Contact Absorber (TCA) system with no forced oxidation.

In addition, preliminary tests have been conducted in the same configurations using adipic acid enhanced lime slurry.

TEST FACILITY AND PROGRAM

There are two scrubber systems operating at the EPA sponsored Shawnee Test Facility, each with its own independent slurry handling facilities. Both systems were tested with adipic acid additive. The systems have the following scrubbers:

- A venturi followed by a spray tower (V/ST)
  (35,000 acfm capacity @ 300°F)
- A Turbulent Contact Absorber (TCA)
  (30,000 acfm capacity @ 300°F)

The scrubbers receive flue gas from TVA Shawnee coal-fired boiler No. 10. The boiler normally burns a high-to-medium sulfur bituminous coal producing SO_2 concentrations of 1500 to 4500 ppm. Flue gas can be taken from either upstream or downstream of the boiler No. 10 particulate removal equipment, allowing testing with high fly ash loadings (3 to 6 grains/scf dry) or low loadings (0.04 to 0.2 grains/ scf dry). All tests in the adipic acid series were made with high fly ash loadings.

The test program was conducted with the scrubbing loop fully closed. Chlorides from the flue gas were concentrated in the scrubber slurry liquor over a range of 1000 to 10,000 ppm depending on the tightness of the scrubber water balance and the chloride concentration in the coal burned.

The Shawnee Test Facility has been operating since March 1972. Bechtel National, Inc. of San Francisco is the major contractor and test director; TVA is the constructor and test facility operator. The initial test program lasted through October 1974[4] with the major emphasis on demonstrating reliable operation. The forced-oxidation tests are a part of an advanced test program that is currently scheduled to continue through December 1979. Earlier results of the advanced test program and a description of the test facility are reported elsewhere.[5,6,7,8,9,10]

The Advanced Test Program schedule for the period covered in this report is shown in Figure 1. As can be seen, testing with adipic acid additive has constituted the major effort during this period.

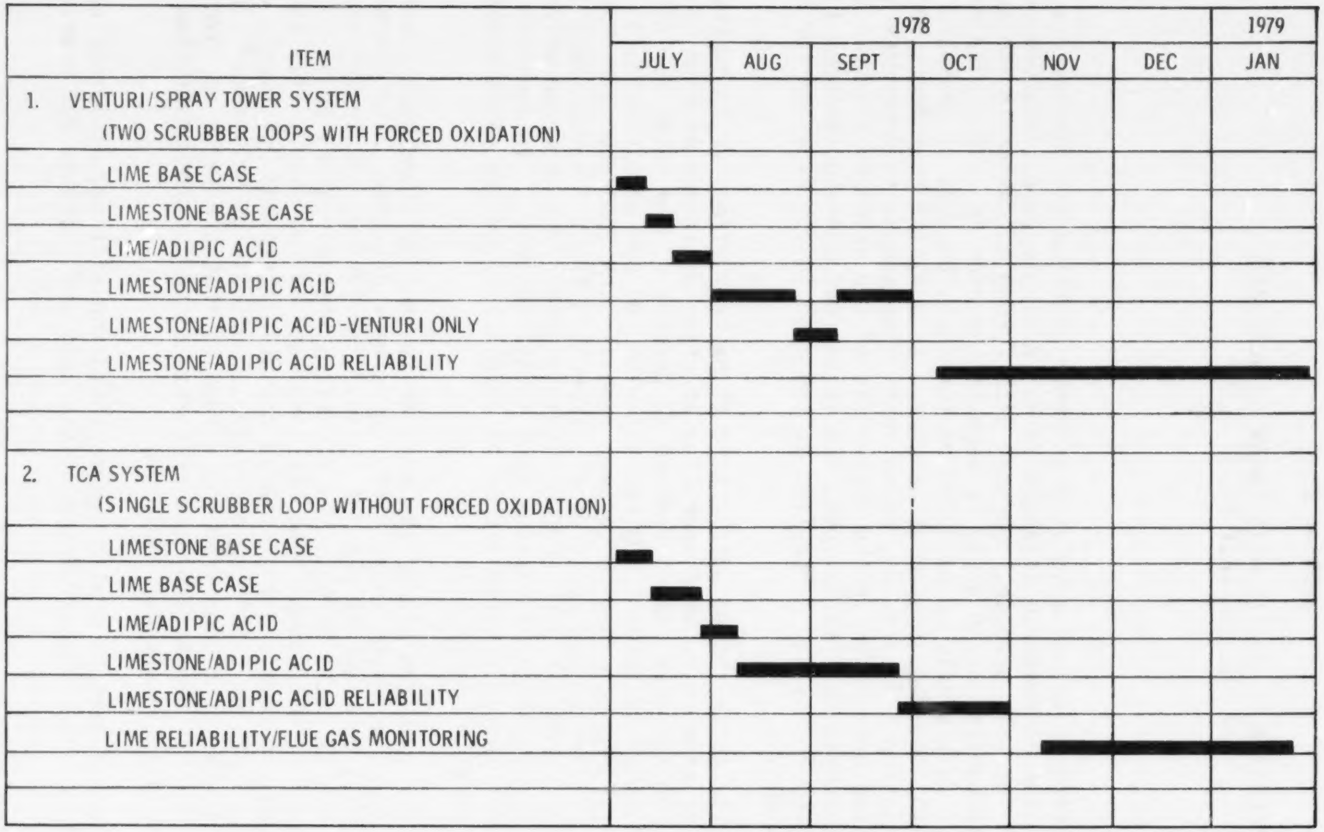

Figure 1. SHAWNEE ADVANCED PROGRAM TEST SCHEDULE FOR PERIOD JULY 1978 THROUGH JANUARY 1979

Section 2

EVALUATION OF ADIPIC ACID ENHANCED SCRUBBING AT THE IERL-RTP PILOT PLANT

The initial testing of adipic acid as a scrubber additive was carried out in the EPA pilot plant at Research Triangle Park. A single-loop limestone scrubber[11] was used for this purpose, operated with forced oxidation in the scrubbing loop. This configuration would be expected to provide a sensitive response to any effect of adipic acid on oxidizer performance, because it operated at a higher pH than the two-loop system at Shawnee. In addition to effects on SO_2 removal and oxidation efficiencies, these tests also sought to determine whether adipic acid caused any change in the properties of the oxidized sludge. So that these properties could be clearly seen, the system was operated without fly ash. Chloride was added as HCl and controlled at the high levels expected for tightly closed-loop systems.

The results of the tests showed adipic acid to be very effective in improving SO_2 removal efficiency, even when operating at chloride levels as high as 17,000 ppm. A TCA scrubber, which removed 82 percent of the inlet SO_2 without the additive, yielded 89 percent SO_2 removal with 700 ppm adipic acid, 91 percent removal with 1000 ppm, and 93 percent removal with 2000 ppm adipic acid. The limestone utilization was concurrently increased from 77 percent without the additive, to 91 percent with 1600 ppm adipic acid. The observed effects thus confirmed the theoretical expectations in all respects. In addition, the tests showed no serious interference by adipic acid on the performance of the oxidizer, operating at pH 6.1.

The quality of the oxidized sludge was similar to that obtained when operating without adipic acid, although small differences were detected. For example, the filtered sludge averaged 80 percent solids (for 13 one-week tests) vs. 84 percent solids for 11 tests without the additive, when operating at 97 to 99 percent oxidation in both cases. The settling rate of the slurry (fly ash free at 50°C) averaged 2.3 cm/min during the adipic acid tests and 3.4 cm/min without adipic acid; bulk settled densities averaged 1.0 and 1.2 gm solids/cm^3 slurry, respectively. It was concluded from these results that the large improvements in sludge quality that can be achieved by forced oxidation are unaffected by the use of adipic acid as a scrubber additive.

Tests without forced oxidation also demonstrated the efficacy of adipic acid. Operating a TCA scrubber with 2000 ppm adipic acid and 6 inches H_2O pressure drop, 92 percent SO_2 removal was obtained at a limestone utilization level of 88 percent. By comparison, only 75 percent SO_2 removal would be expected in the pilot plant at these test conditions without the additive. At this adipic acid level the unoxidized sludge filtered to 49 percent solids; at lower adipic acid levels (1500 ppm or less) the filterability of the slurry was the same as that obtained without additives: 55 percent solids.

During all tests with adipic acid, the scrubbing liquor had a noticeable odor even though the additive feed did not. The odor has been identified as valeric acid [$CH_3(CH_2)_3COOH$], which is an intermediate product formed by side reactions that degrade adipic acid at scrubber operating conditions. Tests conducted by Radian Corporation(12) at IERL-RTP showed 40-50 ppm valeric acid in the scrubbing liquor, and about 1 ppm in the effluent flue gas when operating without forced oxidation and a 2000 ppm adipic acid level. Although laboratory tests show that adipic acid ultimately degrades to lower molecular weight (C_1 to C_4) paraffinic hydrocarbons, no degradation products other than valeric acid have been detected (detection threshold = 10 ppb) in the pilot plant effluents. Efforts by Radian to identify the chemical mechanism responsible for degradation are continuing.

Material balances were carried out with and without forced oxidation to compare the adipic acid makeup requirements. The results showed that the degradation is greater when operating with forced oxidation, in which case 64 percent of the feed was unaccounted for. Without forced oxidation, the estimated loss was 24 percent. In spite of the greater rate of degradation with forced oxidation, the pilot plant material balance indicated that the adipic acid makeup should still be minimal for this mode: the reduction in liquor loss resulting from improved slurry dewatering properties more than compensates for the additional degradation.

Section 3

ADIPIC ACID ENHANCED SCRUBBING IN THE SHAWNEE VENTURI/SPRAY TOWER SYSTEM WITH TWO SCRUBBER LOOPS AND FORCED OXIDATION

Since January 1977, the venturi/spray tower system has operated with two scrubber loops in series and with forced oxidation accomplished by sparging air into the first scrubber loop (venturi) hold tank. Successful demonstration of this mode of scrubbing has already been reported with three alkali types: limestone, lime, and limestone with added magnesium oxide.9,10)

Since July 1978, tests have been conducted with added adipic acid in both lime and limestone systems. With adipic acid, removals as high as 98 percent have been achieved with both alkalis while concurrently oxidizing the product to gypsum. Results of these adipic acid enhanced, forced-oxidation tests are reported in this section.

SYSTEM DESCRIPTION

The venturi/spray tower system was modified for two-loop scrubber operation with forced oxidation as shown in Figure 2. To separate the venturi and spray tower scrubber loops, a catch funnel was installed beneath the bottom spray header of the spray tower. To minimize slurry entrainment through the catch funnel, the bottom spray header was turned upward.

The hold tank in the first scrubber (venturi) recirculation loop was used as the oxidation tank. The arrangement of this tank is shown in Figure 3. The tank was 8 ft in diameter and could be operated at 10, 14, or 18-ft slurry levels. In early tests the tank contained an air sparger ring made of straight 3-inch 316L SS pipe pieces welded into an octagon approximately 4 ft in diameter. It was located 6 inches from the bottom of the tank. Sparger rings had either 130 1/8-inch diameter holes or 40 1/4-inch diameter holes pointing downward. The sparger ring was fed with compressed air to which sufficient water was added to assure humidification. (Dry air can evaporate water at the sparger orifice and cause scaling). In more recent tests the sparger ring was replaced by a 3-inch diameter pipe with an open elbow discharging air downward at the center of the tank about 3 inches from the tank bottom. In all tests with adipic acid enhancement, the 3-inch pipe was used for air discharge.

The oxidation tank had an agitator with two axial flow turbines, both pumping downward. Each turbine was 52 inches in diameter and contained 4 blades. The bottom turbine was 10 inches above the air sparger. The agitator rotated at 56 rpm and was rated at 17 brake Hp.

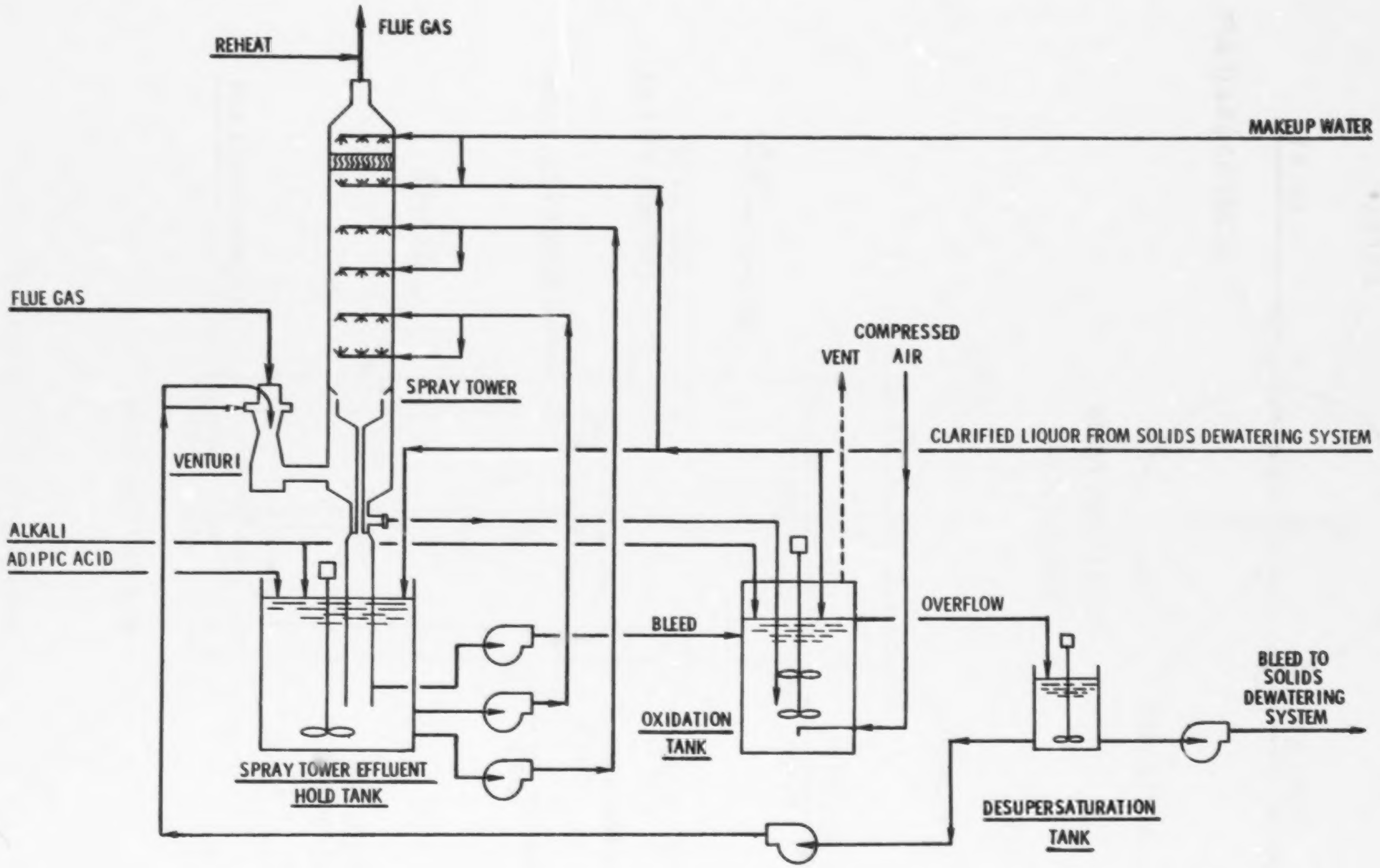
FLUE GAS
REHEAT
MAKEUP WATER
FLUE GAS
COMPRESSED
VENT
AIR
SPRAY TOWER
CLARIFIED LIQUOR FROM SOLIDS DEWATERING SYSTEM
VENTURI
ALKALI
ADIPIC ACID
OVERFLOW
BLEED
BLEED TO SOLIDS DEWATERING SYSTEM
OXIDATION TANK
SPRAY TOWER EFFLUENT HOLD TANK
DESUPERSATURATION TANK

Figure 2. FLOW DIAGRAM FOR ADIPIC ACID ENHANCED SCRUBBING IN THE VENTURI/SPRAY TOWER SYSTEM WITH TWO SCRUBBER LOOPS AND FORCED OXIDATION

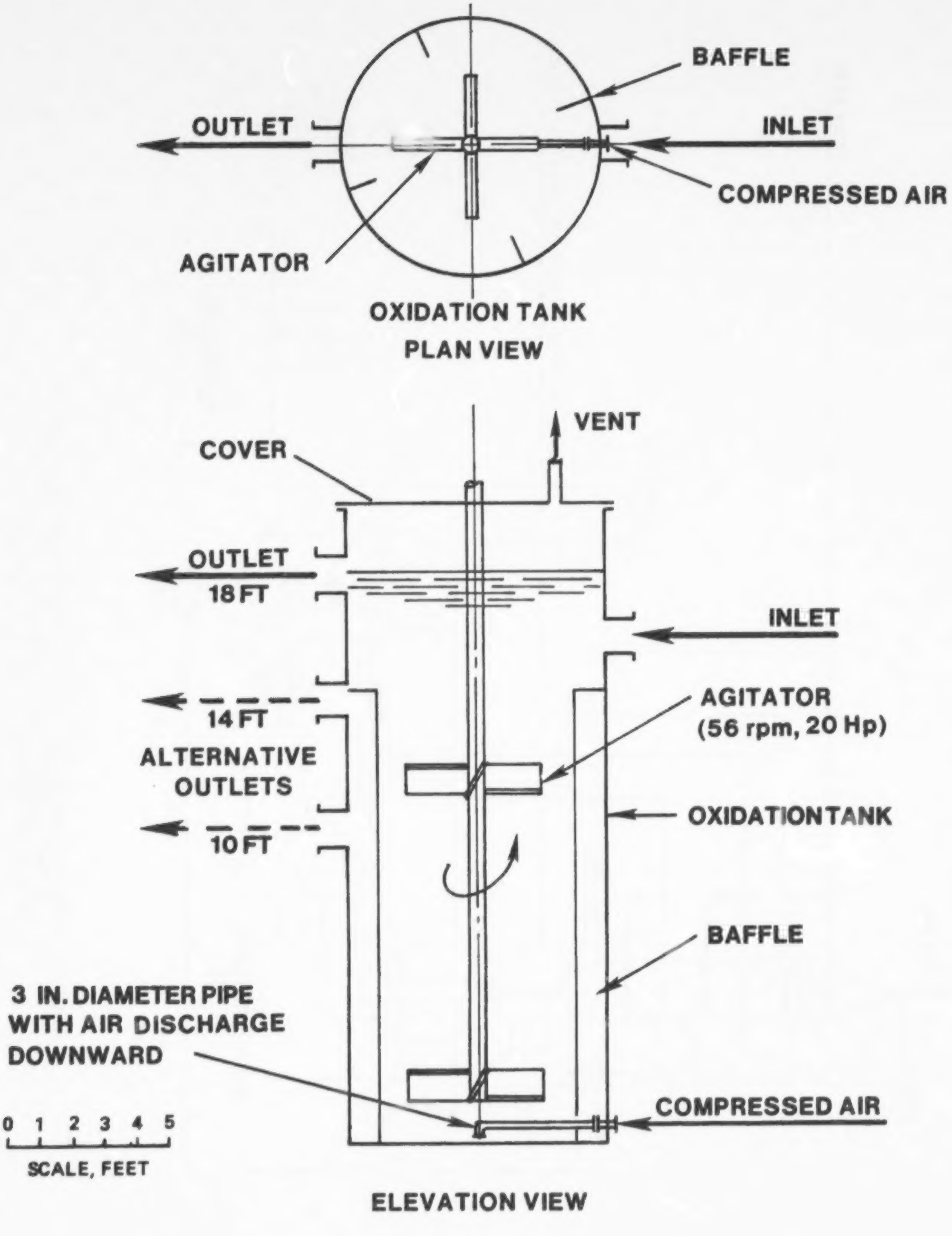
BAFFLE
OUTLET
INLET
COMPRESSED AIR
AGITATOR
OXIDATION TANK
PLAN VIEW
VENT
COVER
OUTLET
18 FT
INLET
14 FT
ALTERNATIVE
OUTLETS
10 FT
AGITATOR
(56 rpm, 20 Hp)
OXIDATION TANK
BAFFLE
3 IN. DIAMETER PIPE
WITH AIR DISCHARGE
DOWNWARD
COMPRESSED AIR
0 1 2 3 4 5
SCALE, FEET
ELEVATION VIEW

Figure 3. ARRANGEMENT OF THE VENTURI/SPRAY TOWER OXIDATION TANK

A 10-ft diameter desupersaturation tank, operating at a 5-ft slurry level, followed the oxidation tank to provide time for gypsum precipitation and to provide air-free pump suction.

Provision was made to add alkali to either loop. Adipic acid was added as a dry powder to the spray tower effluent hold tank. The dewatering system consisted of a clarifier followed by a rotary drum vacuum filter. Clarified liquor from the dewatering system can be returned to either scrubber loop or to the mist eliminator wash circuit.

SUMMARY OF PREVIOUSLY REPORTED TWO-LOOP FORCED-OXIDATION TEST RESULTS

Forced-oxidation test results with two scrubber loops conducted from January 1977 through June 1978 with lime and limestone slurry have been previously reported.[9,10]

Key results of these earlier tests were:

- Oxidation of sulfite solids to gypsum of 90 percent or better dramatically improved the dewatering and handling characteristics of the waste solids.
- Slurry oxidation of better than 96 percent in the first of two independent scrubbing loops was demonstrated with simple air sparging through a 3-inch pipe into an agitated tank with the configuration shown in Figure 3.
- Conditions under which near complete oxidation was demonstrated were an oxidation tank pH range of 4.5 to 5.5, an air stoichiometric ratio of at least 1.5 atoms oxygen/mole of SO_2 absorbed, and an oxidation tank level of at least 14 feet.
- Slurries with high or low fly ash loadings oxidized equally well.
- A slurry solids concentration of 7 percent or higher in the spray tower was required to prevent calcium sulfite scaling and to maintain good SO_2 removal.
- For pH control in both scrubber loops, it was necessary to add lime to both loops. With limestone, addition to the spray tower loop was sufficient.
- In a 32 day run with lime slurry, an average SO_2 removal of 88 percent (2950 ppm average inlet SO_2 concentration) was achieved. Lime utilization was 98 percent.
- In a 35 day run with limestone slurry, an average SO_2 removal of 86 percent (2950 ppm average inlet SO_2 concentration) was achieved. Limestone utilization was 81 percent.

- In a 20 day limestone run with 5000 ppm effective Mg^{++} concentration* in the spray tower scrubber loop, an average SO_2 removal of 96 percent (2250 ppm average inlet SO_2 concentration) was achieved. Limestone utilization was 92 percent.

TWO-LOOP FORCED-OXIDATION TEST RESULTS USING LIMESTONE SLURRY WITH ADDED ADIPIC ACID

Beginning in July 1978, a series of 9 limestone tests with forced oxidation on the two-loop venturi/spray tower system were conducted to demonstrate adipic acid as an additive to enhance SO_2 removal. Results of these tests are summarized in Table 1. Unless otherwise specified, controlled run conditions common to all of the tests were:

Fly ash loading: High (3 to 6 grains/scf dry)
Flue gas rate: 35,000 acfm at 300°F
Venturi slurry rate: 600 gpm (21 gal/Mcf)
Spray tower slurry rate: 1600 gpm (57 gal/Mcf)
Venturi slurry solids concentration: 15 percent
Adipic acid in spray tower liquor: 1500 ppm

After operating variables were explored in runs lasting about a week each, SO_2 removals above 95 percent (2500 ppm inlet SO_2 concentration) were achieved in long term tests at limestone utilizations of about 90 percent and with near complete oxidation.

Effect of Adipic Acid Concentration - The effect of adipic acid on enhancing SO_2 removal can be seen by comparing Run 901-1A (no adipic acid) with Run 902-1A (nominally 1500 ppm adipic acid in the spray tower slurry). Both runs were purposely made at high limestone utilization (97 and 94 percent in Runs 901-1A and 902-1A, respectively) to demonstrate the effect of adipic acid. The addition of adipic acid increased SO_2 removal from 57 percent with no acid to 91 percent at nominally 1500 ppm. These runs were made before conditions for operating with adipic acid were optimized. After optimization, SO_2 removals in excess of 95 percent were routinely achieved at 1500 ppm adipic acid.

Actual adipic acid concentrations were 1445 ppm in the spray tower loop and 2045 ppm in the venturi loop. Dissolved solids are concentrated in the venturi loop because water is evaporated in humidifying the hot flue gas entering the scrubber.

Results at the IERL-RTP pilot plant indicated that most of the enhancement effect with adipic acid may be achieved at concentrations as low as 700 ppm in the slurry liquor. Adipic acid concentrations lower than the 1500 ppm already tested will be explored at the Shawnee facility as time permits.

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* Effective Mg^{++} concentration is defined as the total magnesium ion concentration minus that magnesium ion equivalent to total chlorides. Magnesium chloride has no effect on SO_2 removal.

Table 1

VENTURI/SPRAY TOWER TWO-LOOP TESTS WITH FORCED OXIDATION
- ADIPIC ACID ENHANCED LIMESTONE SLURRY -

Major Test Conditions[1]	901-1A	902-1A	902-1B	903-1A	904-1A	905-1A	906-1A	907-1A[7]	907-1B[7]
Fly ash loading	High	High	High	High	High	High	High	High	High
Adipic acid concentration in venturi, ppm	0	(~2000)	(~2000)	3500	3500	(~2000)	(~2000)	(~2500)	(~2500)
Adipic acid concentration in ST (controlled), ppm	0	1500	1500	-	-	1500[4]	1500	1500	1500
Gas rate, acfm @ 300°F	35,000	35,000	35,000	35,000	18,000	35,000	35,000	18,000-35,000	20,000-35,000
Venturi slurry rate, gpm	600	600	600	600	600	600	600	600	600
ST slurry rate, gpm	1600	1600	1600[2]	0[3]	0[3]	1600	1600	1600	1600
Venturi solids recirculated (controlled), wt.%	15	15	15	15	15	15	15	15	15
Residence times (min)/tank level (ft): Spray tower EHT	14.7/10	14.7/10	14.7/10	-	-	14.7/10	14.7/10	14.7/10	14.7/10
Oxidation tank	11.3/18	11.3/18	11.3/18	11.3/18	11.3/18	11.3/18	11.3/18	11.3/18	11.3/18
Desupersaturation tank	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8	4.7/4.8
Venturi inlet pH (estimate, not controlled)	(~4.5)	(~4.5)	(~4.5)	(~5.5)	(~5.5)	4.5[5]	4.5[5]	4.5[5]	4.5[5]
Venturi limestone stoichiometric ratio (controlled)	-	-	-	1.3	1.3	-	-	-	-
ST limestone stoichiometric ratio (controlled)	1.4	1.4	1.4	-	-	1.45	1.70	1.70	1.70
Air rate to oxidizer, scfm	210	260	260	210	130	260	260	260	260
Run Average Results									
Start-of-Run date	7/12/78	8/3/78	8/22/78	8/28/78	9/5/78	9/8/78	9/27/78	10/8/78	11/13/78
Onstream hours	107	373	65	183	72	439	153	719	1666
SO_2 removal, %	57	91	89	53	67	86	93	97.5	97
Inlet SO_2 concentration, ppm	2800	2450	2400	2600	2350	2200	2700	2350	2500
Spray tower solids recirculated, wt.%	5.9	4.9	4.8	-	-	3.2	6.8[6]	6.1[6]	5.9[6]
Venturi inlet pH	4.50	4.45	4.00	4.7	4.75	4.55	4.5	4.65	4.65
Spray tower inlet pH	5.45	5.25	5.05	-	-	4.75	5.15	5.45	5.35
Spray tower limestone stoichiometric ratio	1.36	1.65	1.38	-	-	1.53	1.49	1.77	1.70
Spray tower inlet liquor gypsum saturation, %	95	110	115	-	-	130	100	105	110
Spray tower sulfite oxidation, %	28	40	27	-	-	63	28	29	25
Overall sulfite oxidation, %	98	98	99	98	98	99	98.5	98.5	98
Overall limestone utilization, %	97	94	98	74	76	91	88	88	92
Venturi inlet liquor gypsum saturation, %	95	100	90	100	100	100	110	110	105
Venturi inlet liquor $SO_3^=$ concentration, ppm	80	55	45	65	60	60	70	65	75
Adipic acid concentration in venturi, ppm	0	2045	2355	3510	3690	2315	2495	2360	2180
Adipic acid concentration in ST, ppm	0	1445	1615	-	-	1410	1485	1560	1510
Air stoichiometry, atoms O/mole SO_2 absorbed	2.30	2.05	2.15	2.65	2.80	2.40	1.80	2.0-3.85	1.9-3.3
Filter cake solids, wt%	85	86	88	83	80	85	84	87	85
Mist eliminator restriction, %	0.2	0.2	-	-	0	0	0	0	-

Notes:
(1) All runs were made with 9 in. H_2O venturi pressure drop except Runs 907-1A and 907-1B in which pressure drop was 9 in. H_2O at 35,000 acfm.
(2) Intermittent spray tower operation. Spray tower slurry flow turned off 30 minutes every 8 hours. SO_2 removal averaged ~30 percent by venturi only operation.
(3) Venturi alone operation. The spray tower slurry recirculation pumps were turned off for the entire run.
(4) During the initial portion of the run (unsteady state operation) adipic acid was allowed to deplete to observe SO_2 removal/adipic acid relationship.
(5) Venturi inlet pH was controlled by separate limestone addition.
(6) Clarified liquor in excess of mist eliminator wash was returned to spray tower EHT, except for Runs 906-1A, 907-1A, and 907-1B where a fraction of this stream was diverted to the oxidation tank to control spray tower slurry solids at 6%.
(7) Long-term reliability test.

SO_2 Removal in the Venturi - Three runs were made to determine SO_2 removal in the venturi alone. In Run 902-1B the spray tower recirculation pumps were turned off for 30 minutes every 8 hours. In Runs 903-1A and 904-1A, only the venturi was operated. Results were:

V/ST Run	Flue Gas Rate, acfm	Adipic Acid in Venturi, ppm	Venturi Inlet pH	% SO_2 Removal in Venturi
902-1B	35,000	2355	4.0	30
903-1A	35,000	3510	4.7	53
904-1A	18,000	3690	4.8	67

In Run 902-1B, SO_2 removal was no greater than usually achieved without adipic acid. This is not unexpected because the operating pH was below the 4.0 to 5.0 buffering range of adipic acid.

In Run 903-1A, the pH was increased to 4.7 (limestone stoichiometric ratio of 1.35 mole Ca /mole SO_2 removed) and adipic acid concentration was increased to nominally 3500 ppm with a resultant increase in SO_2 removal to 53 percent. Finally in Run 904-1A, the flue gas flow rate was cut in half to double the liquid-to-gas ratio, resulting in a further increase in SO_2 removal to 67 percent. These tests showed that high removal efficiencies cannot be achieved with reasonably low limestone stoichiometry in the venturi alone. Although not yet demonstrated, it may be possible to increase pH and get high removal efficiency using lime with adipic acid in the venturi alone.

System Control - In previous limestone runs without adipic acid, the limestone stoichiometry in the spray tower loop was controlled by adding limestone to the spray tower hold tank. By using a spray tower limestone stoichiometric ratio of about 1.4, there was sufficient residual limestone in the bleed from the spray tower loop to the venturi loop to maintain the venturi inlet slurry liquor pH above 5.0. This was not the case with adipic acid addition. With adipic acid, the venturi inlet pH fluctuated between 4.0 and 4.5 which was too low to get good removal.

Beginning with Run 905-1A, limestone was added to both the spray tower and venturi loops in an attempt to raise the venturi inlet pH. The pH could not be raised above 4.9 even with a venturi limestone stoichiometry greater than 2.0. For reasons not yet clear, adipic acid has a depressing effect on the pH in the venturi loop.

In subsequent runs, the spray tower limestone stoichiometric ratio was increased to 1.7 to maintain a venturi inlet pH of 4.5. This pH was established as the highest pH in which an overall limestone utilization of about 90 percent could be achieved. Occasionally, limestone is still added to the venturi loop if the venturi inlet pH drops much below 4.5.

Effect of Slurry Solids Concentration - In unenhanced limestone systems, slurry liquor pH and, consequently, SO_2 removal are sensitive to slurry solids concentration (i.e., limestone surface area available for dissolution) below 8 percent. Because of increased liquor buffer capacity with adipic acid enhancement, the system should be less sensitive to solids concentration. However, in early runs with adipic acid, the slurry solids concentration was allowed to drop to sensitive levels. In these runs, clarified liquor from the thickener was returned to the spray tower loop which resulted in low solids concentrations in the spray tower slurry in the range of 3 to 5 percent.

Beginning with Run 906-1A, the spray tower solids concentration was controlled at nominally 6 percent by splitting the clarified liquor return between the spray tower and the venturi loops. The spray tower pH and, consequently, SO_2 removal were increased as can be seen by the following comparison:

V/ST Run	Spray Tower % Solids	Spray Tower Inlet pH	Spray Tower Limestone Stoich.	Inlet SO_2 Conc. ppm	Percent SO_2 Removal
905-1A	3.2	4.75	1.5	2200	86
906-1A	6.8	5.15	1.5	2700	93

Based on an examination of day-to-day operations, with 1500 ppm adipic acid in limestone slurry, it appears that SO_2 removal increases with spray tower slurry solids concentration up to about 6 weight percent.

Forced Oxidation - Forced oxidation of the scrubber slurry has been shown to occur equally well with or without adipic acid. In fact, adipic acid buffers the slurry liquor at a low pH favorable for forced oxidation. The adipic acid enhanced runs on the two-loop venturi/spray tower system were all operated at air stoichiometric ratios in the range of 1.8 to 2.4 atoms oxygen/mole SO_2 absorbed. Near complete oxidation was achieved in all runs and the filter cake solids concentration of the oxidized slurry was consistently high, in the range of 85 percent or better.

Venturi/Spray Tower Demonstration Runs 907-1A and 907-1B - Run 907-1A was a month long adipic acid enhanced limestone run with forced oxidation designed to demonstrate operational reliability with respect to scaling and plugging and to demonstrate the removal enhancement capability of the adipic acid additive.

This run was controlled at a nominal 1.7 limestone stoichiometric ratio (up from 1.4 in previous runs) and 1500 ppm adipic acid in the spray tower. Spray tower slurry solids concentration was controlled at 6 percent by splitting the clarified liquor return between the spray tower and venturi loops. Venturi inlet slurry liquor pH was nominally 4.5. Occasionally, limestone addition to the venturi loop was required to maintain this pH.

Flue gas flow rate was varied from 18,000 acfm to a maximum of 35,000 acfm (spray tower gas velocity between 4.8 and 9.4 ft/sec) to follow the daily boiler load cycle which normally fluctuated between 100 and 150 MW. The adjustable venturi plug was fixed in a position such that the pressure drop across the venturi was 9 inches H_2O at 35,000 acfm maximum gas rate. Actual pressure drop ranged from 3.0 to 9.6 inches H_2O.

The slurry recirculation rates to the venturi and spray tower were fixed at 600 gpm (L/G = 21 to 42 gal/Mcf) and 1600 gpm (L/G = 57 to 111 gal/Mcf), respectively.

The oxidation tank level was 18 ft and the air flow rate was held constant at 260 scfm.

The run began on October 8, 1978 and terminated November 13, 1978. It ran for 719 on-stream hours (30 days) with no unscheduled outages. The scrubber was down once for a scheduled 3-hour inspection and again when the boiler came down for 135 hours to install a new station power transformer.

Average SO_2 removal for the run was 97.5 percent at 2350 ppm average inlet SO_2 concentration. The SO_2 removal stayed within a narrow range of 96 to 99 percent throughout almost the entire run. On October 19 and on October 27, SO_2 removal dropped briefly to less than 90 percent when the pump which supplies the slurry to the top two spray headers was brought off stream for repacking and the spray tower slurry flow rate was cut in half to 800 gpm. At the reduced slurry recirculation rate, SO_2 removal was 82 to 87 percent.

Venturi and spray tower inlet pH averaged 4.65 and 5.45, respectively. Overall limestone utilization was 88 percent and the spray tower limestone utilization was 56 percent, demonstrating the advantage of good limestone utilization in a two-scrubber-loop operation.

Average adipic acid concentrations were 2360 ppm in the venturi loop and 1560 ppm in the spray tower loop.

Sulfite oxidation in the system bleed slurry averaged 98.5 percent with the air stoichiometric ratio varied between 2.0 and 3.85 atoms oxygen/mole SO_2 absorbed. The filter cake solids content was 87 percent.

The mist eliminator was clean during the entire run. Inspections after 207 hours, 603 hours, and 719 hours (at the end of the run) showed that the mist eliminator was completely free of solids deposits. The system was free of plugging and scaling. There was no increase in solids or scale deposits on the scrubber internals during Run 907-1A.

Following Run 907-1A, a second adipic acid enhanced limestone run with forced oxidation was made during which flue gas monitoring procedures were evaluated by EPA. This run, 907-1B, was made under the same conditions as Run 907-1A except that a "typical" daily boiler load cycle was established for the gas flowrate to follow rather than the Unit No. 10 Boiler load. The gas rate was changed as follows to simulate a "typical" boiler load cycle:

Time, hours	Gas Rate, acfm @ 300°F
0100	20,000
0500	30,000
0700	35,000
1100	30,000
1700	35,000
2300	30,000

Run 907-1B began on November 13, 1978 and terminated January 29, 1979. It ran for 1666 on-stream hours (69 days) with only 27 hours of scrubber related outages. The scrubber was also out of service 146 hours when Unit 10 came down for replacement of a broken turbine thrust bearing. Scrubber related outages were:

Plugged slurry line	11 hours
Rebuild bleed pump	10 hours
Miscellaneous mechanical	4 hours
Leak in water supply line	2 hours
Total	27 hours

Excluding boiler outages and scheduled inspections, the combined Runs 907-1A and 907-1B operated for a period of over 3 months with an on-stream factor of 98.9 percent. No deposits whatsoever were observed in the mist eliminator for the entire 3 month test period. On only one occasion did solids accumulation cause an outage. The cross-over line carrying slurry effluent from the venturi to the oxidation tank plugged with soft solids and had to be cleaned out. Because of problems associated with converting the Shawnee venturi/spray tower system to two-scrubber-loop operation, this cross-over line follows a tortuous path (see Figure 2). A properly designed new system would not have this problem.

Results of Run 907-1B were equally as good in every respect as those of Run 907-1A. Average SO_2 removal was 97 percent at 2500 ppm average inlet SO_2. With few exceptions, SO_2 removal remained within a narrow band of 95 to 99 percent. SO_2 removal dropped briefly (typically 30 minutes) below 90 percent five times when one of the two spray tower recirculation pumps was taken out of service for maintenance, effectively cutting the slurry recirculation rate in half.

Overall limestone utilization during this run was 92 percent. Sulfite oxidation averaged 98 percent and the waste sludge filter cake quality was excellent, having a solids content of 85 percent.

SO_2 emissions for Run 907-1A and 907-1B were calculated based on an assumed coal heating value of 10,500 Btu/lb, on 100 percent sulfur overhead (none in bottom ash), and on an assumed excess air of 30 percent. This excess air corresponds to about 700 ppm inlet SO_2 per 1.0 weight percent sulfur in coal for the above conditions.

Figures 4 and 5 show calculated SO_2 emissions for Runs 907-1A and 907-1B, respectively. Average SO_2 emissions for each 24-hour period (midnight-to-midnight) are indicated by horizontal lines on the figures. The average SO_2 emission for the entire 3-month operating period was only 0.20 lb/MM Btu. The highest 24-hour average SO_2 emission during Run 907-1A was 0.37 lb/MM Btu (October 18) and during Run 907-1B was 0.41 lb/MM Btu (January 20). These values compare with the federal new source performance standard of 1.2 lb/MM Btu.

Adipic Acid Consumption - From the onset of testing, adipic acid consumption was higher than anticipated. A material balance calculation for the adipic acid consumption was made for the entire Run 907-1B for a period of 76 days from November 13, 1978 to January 29, 1979. During this period, a breakdown of the average adipic acid addition rate was:

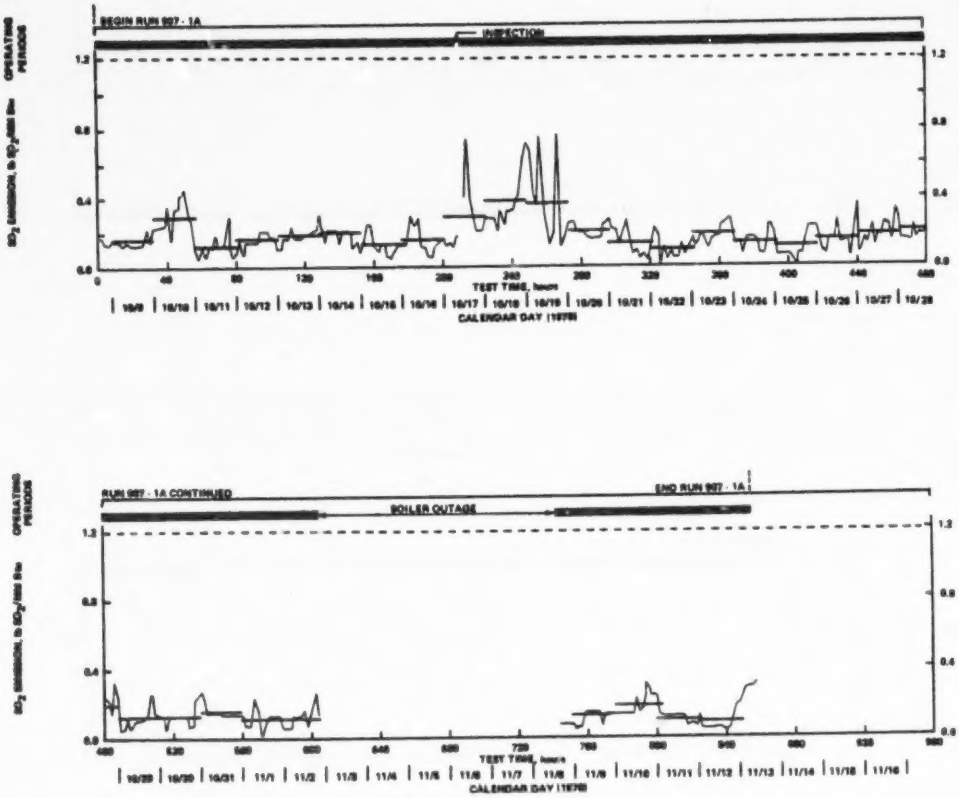
Run 907-1B Adipic Acid Consumption

	lb/hr	lb/ton limestone feed
Discharged with filter cake	0.8	1.8
Unaccounted	2.9	6.5
Total	3.7	8.3

These losses were higher than experienced in the IERL-RTP pilot plant. The reasons for the differences are not yet clear.

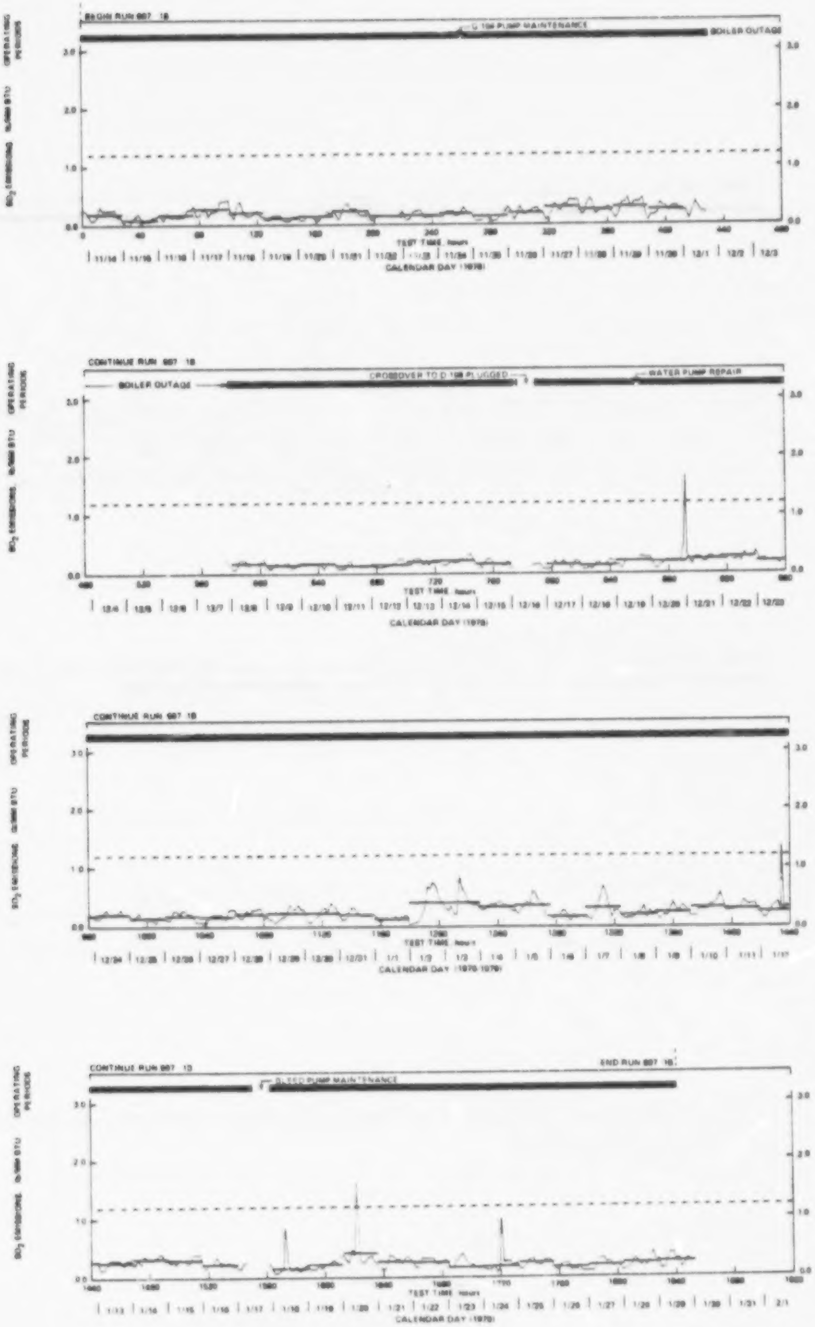
The unaccounted adipic acid loss in this run with forced oxidation was somewhat higher than that recorded in TCA Run 932-2A without forced oxidation (2.2 lb/hr, see Section 4). Based on this comparison, it appears that forced oxidation in the scrubber system may increase the unaccounted adipic acid losses.

As already mentioned in Section 2, Radian Corporation is investigating the mechanism of this unaccounted loss. Preliminary results indicate that adipic acid decomposes to valeric acid [$CH_3(CH_2)_3COOH$] and other components. Valeric acid creates an odor even in the small concentrations present in the scrubber slurry. An unpleasant odor was apparent immediately above the effluent hold tanks and in the filter and centrifuge building where dewatering takes place. About 600 cubic yards of the gypsum/fly ash mixture from the 3-month combined venturi/



Note: Horizontal bars indicate 24-hour (midnight-to-midnight) run averages

Figure 4. SO_2 EMISSIONS DURING VENTURI/SPRAY TOWER RUN 907-1A



Note: Horizontal bars indicate 24-hour (midnight-to-midnight) run averages

Figure 5. SO_2 EMISSIONS DURING VENTURI/SPRAY TOWER RUN 907-1B spray tower Runs 907-1A and 907-1B were saved in a pile near the test facility. During this winter no odor was apparent even when the pile was worked by a bulldozer. The pile will continue to be monitored during the summer to determine if an odor does exist.

TWO-LOOP FORCED-OXIDATION TEST RESULTS USING LIME SLURRY WITH ADDED ADIPIC ACID

The emphasis at Shawnee has been on adipic acid enhancement with limestone because this combination may prove to be the most economical route for achieving high removal in throwaway systems. However, two one-week runs made in July 1978 demonstrated that adipic acid is also effective in enhancing SO_2 removal in lime systems. Results of these lime runs are summarized in Table 2.

Effect of Adipic Acid Concentration - Two runs were conducted with lime on the venturi/spray tower two-loop system with forced oxidation. Run 951-1A was a base case without adipic acid and Run 952-1A was under identical conditions with nominally 1500 ppm adipic acid in the spray tower slurry liquor. The addition of adipic acid increased SO_2 removal from 66 percent with no acid to 98 percent at 1500 ppm (2600 to 2750 average inlet SO_2 concentration).

Effect of Slurry Solids Concentration - In these lime runs, all clarified liquor from the dewatering system was returned to the spray tower loop resulting in relatively low spray tower solids concentrations. Apparently, a low solids concentration of 4.5 percent in the adipic acid enhanced lime run was not detrimental because 98 percent SO_2 removal was achieved. This may be attributed to the high pH inherent with lime systems, at which adipic acid becomes fully effective. However, in the lime run without adipic acid, the low solids concentration was detrimental. Run 951-1A averaged 66 percent SO_2 removal at 5.7 percent spray tower slurry solids concentration. This run can be compared with a previous one-month lime demonstration run under essentially the same conditions which averaged 88 percent SO_2 removal with 10.4 percent average spray tower slurry solids concentration.

Additional adipic acid enhanced lime tests will be made as time permits.

Table 2

VENTURI/SPRAY TOWER TWO-LOOP TESTS WITH FORCED OXIDATION - ADIPIC ACID ENHANCED LIME SLURRY -

Major Test Conditions(1)	951-1A	952-1A
Fly ash loading	High	High
Adipic acid concentration in venturi, ppm	0	(2000)
Adipic acid concentration in ST (controlled), ppm	0	1500
Gas rate, acfm @ 300°F	35,000	35,000
Venturi slurry rate, gpm	600	600
ST slurry rate, gpm	1600	1600
Venturi solids recirculated (controlled), wt.%	15	15
Residence times (min)/tank level (ft): Spray tower EHT	14.7/10	14.7/10
Oxidation tank	11.3/18	11.3/18
Desupersaturation tank	4.7/4.8	4.7/4.8
Venturi inlet pH (controlled)(2)	5.5	5.5
ST inlet pH (controlled)	7.8	7.8
Air rate to oxidizer, scfm	210	260
Run-Average Results:		
Start-of-Run Date	7/3/78	7/20/78
Onstream hours	250	278
SO_2 removal, %	66	98
Inlet SO_2 concentration, ppm	2750	2600
Spray tower solids recirculated, wt.%	5.7	4.5
Venturi inlet pH	5.6	5.5
Spray tower inlet pH	7.90	7.75
Spray tower lime stoichiometric ratio	1.14	1.23
Spray tower inlet liquor gypsum saturation, %	90	85
Spray tower sulfite oxidation. %	18	22
Overall sulfite oxidation, %	95	97
Overall lime utilization, %	97	98
Venturi inlet liquor gypsum saturation, %	95	90
Venturi inlet liquor $SO_3^=$ concentration, ppm	130	115
Adipic acid concentration in venturi, ppm	0	1585
Adipic acid concentration in ST, ppm	0	1380
Air stoichiometry, atoms O/mole SO_2 absorbed	2.05	1.80
Filter cake solids, wt.%	86	85
Mist eliminator restriction, %	0.5	1

Notes: (1) All runs made with 9 in. H_2O venturi pressure drop. Makeup water and clarified liquor in excess of mist eliminator wash were returned to spray tower EHT.

(2) Venturi inlet pH was controlled by separate lime addition.

Section 4

ADIPIC ACID ENHANCED SCRUBBING IN THE SHAWNEE TCA SYSTEM WITHOUT FORCED OXIDATION

Beginning in July and continuing to November 1978, a series of tests were conducted on the TCA system with adipic acid enhanced lime and limestone slurries. These tests were conducted without forced oxidation. SO_2 removals ranging up to 95 percent have been achieved with adipic acid concentrations of up to 1500 ppm in the recirculating slurry liquor. Results of these adipic acid enhanced tests without forced oxidation are reported in this section.

SYSTEM DESCRIPTION

The TCA (turbulent contact absorber) was operated in a single-loop scrubbing configuration as shown in Figure 6. The TCA contained three beds of 1-7/8 inch diameter, 11.5 gram nitrile foam spheres retained between bar grids. Each bed contained 5 inches static height of spheres.

The effluent hold tank was 7 feet in diameter and operated at a 17 foot slurry level giving a 4.1 minute residence time at the slurry recirculation rate of 1200 gpm. Alkali and adipic acid were added directly to the effluent hold tank.

The dewatering system consisted of a clarifier followed by a solid bowl centrifuge. All clarified liquor from the dewatering system was returned to the scrubber loop either via the effluent hold tank or the mist eliminator underside sprays.

TEST RESULTS ON THE TCA WITH ADIPIC ACID ENHANCED LIMESTONE SCRUBBING

Beginning in July 1978, a series of 7 limestone tests were conducted on the TCA system to demonstrate the effects of adipic acid on enhancing SO_2 removal in a system without forced oxidation. Results of these tests are summarized in Table 3. Unless otherwise specified, controlled run conditions common to all of the tests were:

Fly ash loading: High (3 to 6 grains/scf dry)
Flue gas rate: 30,000 acfm at 300°F
Slurry flow rate: 1200 gpm (50 gal/Mcf)
Slurry solids concentration: 15 percent

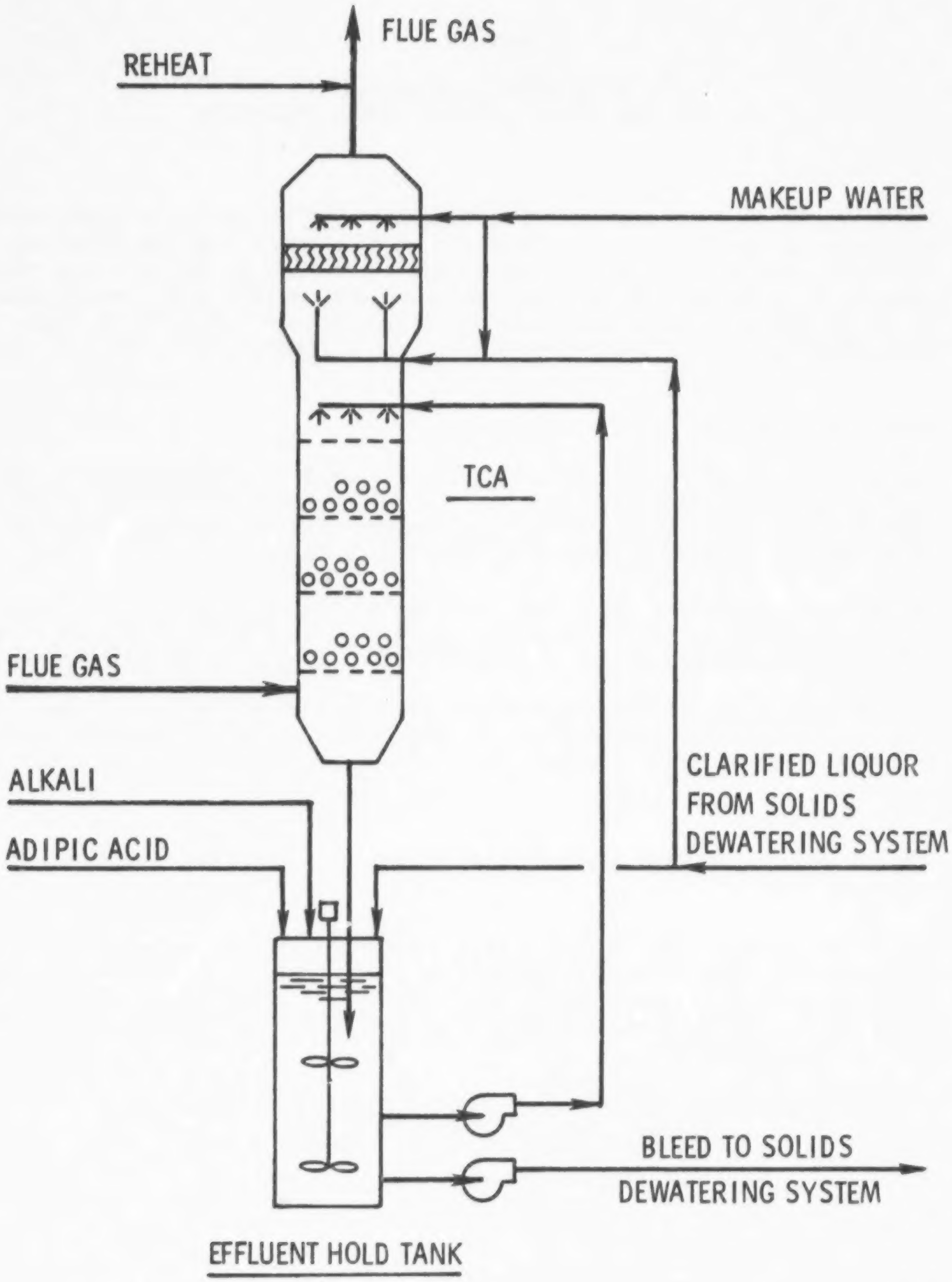

FLUE GAS
REHEAT
MAKEUP WATER
TCA
FLUE GAS
CLARIFIED LIQUOR FROM SOLIDS DEWATERING SYSTEM
ALKALI
ADIPIC ACID
BLEED TO SOLIDS DEWATERING SYSTEM
EFFLUENT HOLD TANK

Figure 6. FLOW DIAGRAM FOR ADIPIC ACID ENHANCED SCRUBBING IN THE TCA SYSTEM

Table 3

TCA SINGLE-LOOP TESTS WITHOUT FORCED OXIDATION
- ADIPIC ACID ENHANCED LIMESTONE SLURRY -

Major Test Conditions[1]	926-2A	927-2A	928-2A	929-2A	930-2A	931-2A	932-2A[3]
Fly ash loading	High	High	High	High	High	High	High
Adipic acid concentration, ppm	0	300	1500	750	750	750	1500
Gas rate, acfm @ 300°F	30,000	30,000	30,000	30,000	30,000	30,000	20,000-30,000
Slurry rate, gpm	1200	1200	1200	1200	1200	1200	1200
Solids recirculated, wt%	15	15	15	15	15	15	15
LS stoichiometric ratio controlled at	1.2	1.2	1.2	1.2	1.35	1.05	1.2
EHT residence time, min.	4.1	4.1	4.1	4.1	4.1	4.1	4.1
EHT level, ft	17	17	17	17	17	17	17
Limestone addition point[2]	EHT	EHT	EHT	EHT	EHT	EHT	EHT
Run Average Results							
Start-of-Run Date	7/4/78	8/4/78	8/22/78	9/5/78	9/13/78	9/20/78	9/26/78
Onstream hours	192	374	252	186	162	137	833
SO_2 removal, wt.%	71	75	93	92	93	77	96
Inlet SO_2 concentration, ppm	2750	2300	2600	2300	2550	2300	2450
SO_2 make-per-pass, m-moles/liter	10.1	9.1	12.8	11.2	12.6	9.4	4-18
TCA inlet liquor gypsum saturation, %	90	110	100	110	80	110	110
Sulfite oxidation, %	13	19	14	10	13	24	21
Limestone utilization, %	80	84	85	80	75	93	82
TCA inlet pH	5.65	5.30	5.40	5.50	5.60	4.95	5.30
Adipic acid concentration, ppm	0	350	1600	885	700	840	1620
Centrifuge solids, wt.%	37[4]	62	59	59	58	62	61
Mist eliminator restriction, percent	1	3	0	0	0	0	0

Notes:
(1) All runs were made with 3 beds and 5 inches per bed of 1-7/8 inch diameter, 11.5 gram nitrile foam spheres.
(2) EHT = in the effluent hold tank.
(3) Long-term reliability test.
(4) Clarifier only.

Effect of Adipic Acid Concentration - TCA limestone Runs 926-2A through 929-2A were all made under identical conditions except for adipic acid concentration. In these runs limestone stoichiometry was controlled at 1.2 moles Ca/mole SO_2 absorbed. The effect of adipic acid on SO removal was:

TCA Run	Actual Adipic Acid Conc., ppm	Percent SO_2 Removal
926-2A	0	71
927-2A	350	75
929-2A	885	92
928-2A	1600	93

SO_2 removal increased from 71 percent with no adipic acid to 92 percent with 885 ppm in the slurry liquor. Increasing adipic acid further to 1600 ppm increased SO_2 removal only slightly to 93 percent. Thus, the majority of the adipic acid enhancement was achieved in the TCA at a concentration somewhere between 350 and 885 ppm.

Effect of Limestone Stoichiometry and pH - Limestone stoichiometry was explored at a nominal adipic acid concentration of 750 ppm and a liquid-to-gas ratio of 50 gal/Mcf with the following results:

TCA Run	Limestone Stoichiometry, mole Ca/mole SO_2 absorbed	TCA Inlet pH	Percent SO_2 Removal
931-2A	1.05	4.9	77
929-2A	1.20	5.5	92
930-2A	1.35	5.6	93

Thus, it is apparent that the system required a limestone stoichiometry of only about 1.2 to maintain sufficiently high pH to achieve the full SO_2 removal enhancement with adipic acid. Additional limestone did not significantly increase SO_2 removal.

TCA Demonstration Run 932-2A - Run 932-2A, a month long demonstration run, was made with adipic acid enhanced limestone slurry to demonstrate both operational reliability with respect to scaling and plugging and the removal enhancement capability of the adipic acid additive.

The run began on September 26, 1978 and terminated on November 2, 1978 for a total of 833 on-stream hours (35 days). During the run, the scrubber was out of service for 48 hours due to a boiler outage caused by a tube leak, 5 hours for a scheduled inspection, and 8 hours for unscheduled outages to clean and repair the scrubber I.D. fan damper.

Excluding boiler outages and scheduled inspections, Run 932-2A operated with an on-stream factor of 99.0 percent.

The run was controlled at a nominal 1.2 limestone stoichiometric ratio and 1500 ppm adipic acid concentration in the slurry liquor. Slurry solids concentration was controlled at 15 percent. The flue gas flow rate was varied between 20,000 and 30,000 acfm (8.4 to 12.5 ft/sec superficial gas velocity) as the boiler load fluctuated between 100 and 150 MW. The slurry recirculation rate was fixed at 1200 gpm (L/G = 50 to 75 gal/Mcf). As with all runs during this test block, the effluent hold tank residence time was only 4.1 minutes.

SO_2 removal during the run averaged 96 percent at an average inlet SO_2 concentration of 2450 ppm. Excluding the first few days of unsteady-state operation, SO_2 removal stayed within the narrow range of 94 to 98 percent as the inlet SO_2 concentration varied widely between 1400 and 3500 ppm.

SO_2 emissions were calculated for Run 932-2A on the same basis as for the venturi/spray tower Run 907-1A (see Section 3). Figure 7 shows the calculated SO_2 emissions for the TCA run. Average SO_2 emissions for each 24-hour period (midnight-to-midnight) are indicated by horizontal lines on the figure.

During the first seven days (September 26 through October 3), SO_2 emissions were relatively high and widely varying. The highest daily average emissions were 1.1 lb SO_2/MM Btu on September 27 and 0.9 lb SO_2/MM Btu on both September 30 and October 3. It should be noted, however, that the new source performance standard of 1.2 lb/MM Btu was never exceeded on a daily average basis.

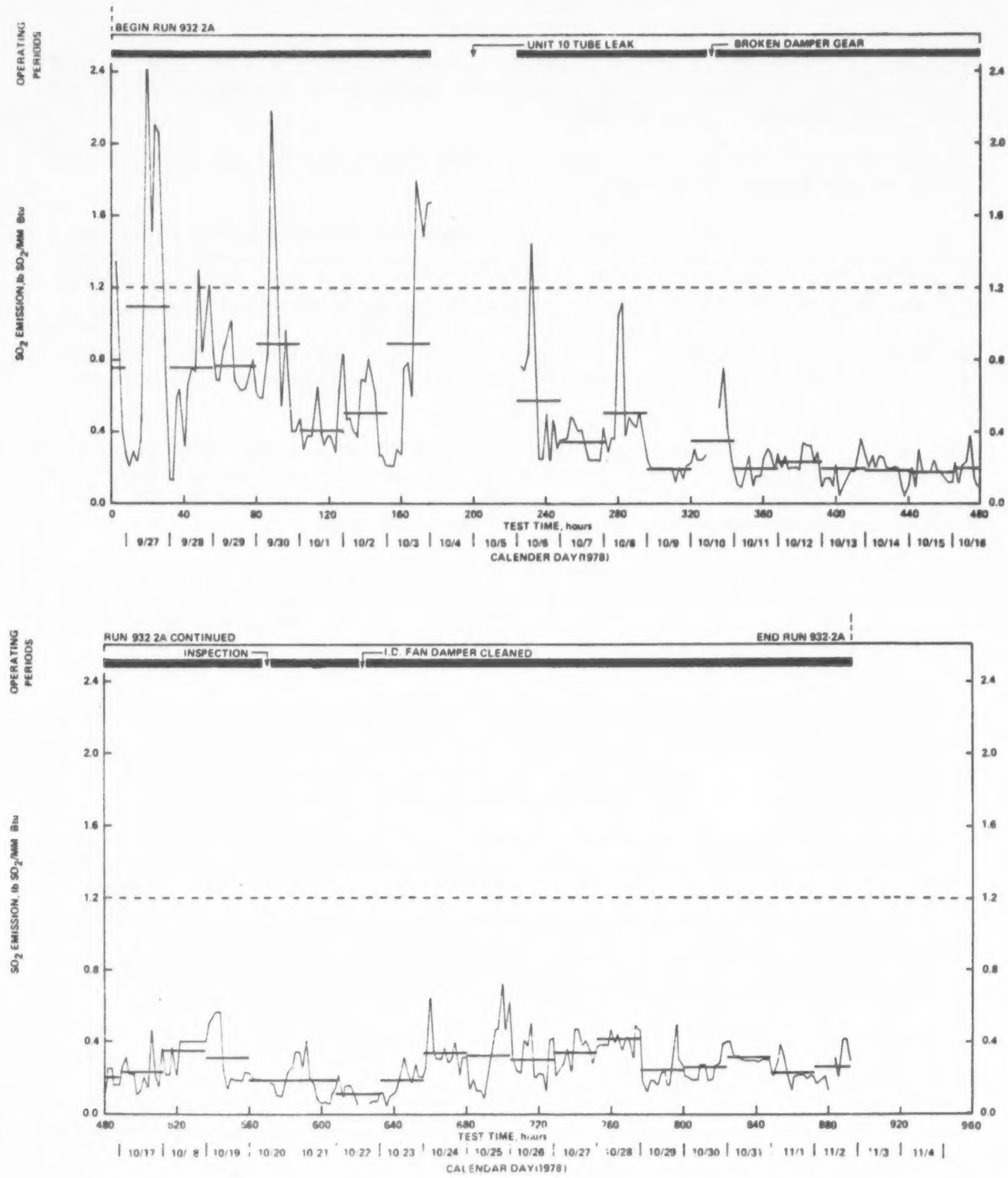
The relatively high SO_2 emissions resulted from frequent excursions to low pH in the scrubber slurry liquor as the test personnel were trying to control the limestone stoichiometric ratio at 1.2 with widely varying inlet SO_2 concentrations (1000 to 3500 ppm). Beginning on October 6 after the boiler outage, a scrubber inlet pH underride of 5.1 was implemented in addition to the limestone stoichiometric ratio control value of 1.2. This combined stoichiometry/pH control produced the improved results shown for the remainder of the run.

SO_2 emissions for the 27-day period from October 6 through the end of the run on November 2 averaged only 0.26 lb/MM Btu. The highest 24-hour average SO_2 emission during this period was only 0.44 lb/MM Btu.

The mist eliminator was completely clean at the end of the run and the entire scrubber system was free of scaling and plugging.

Limestone utilization during the run averaged 82 percent. Discharge solids from the centrifuge averaged about 61 percent which is typical of unoxidized limestone sludge.

In summary, the objectives of this run were met. High removal was consistently achieved at a good limestone utilization and no fouling, scaling, or plugging occurred.



Note: Horizontal bars indicate 24-hour (midnight-to-midnight) run averages

Figure 7. SO_2 EMISSIONS DURING TCA RUN 932-2A

Adipic Acid Consumption - As with the forced-oxidation runs on the venturi/spray tower system, adipic acid consumption was greater than anticipated. An adipic acid material balance calculation was made during Run 932-2A between October 10 and October 30, 1978, a total of 21 days.

Run 932-2A Adipic Acid Consumption

	lb/hr	lb/ton limestone feed
Discharged with centrifuge cake	1.9	4.2
Unaccounted	2.2	5.0
Total	4.1	9.2

As already discussed in Section 3, the unaccounted loss in this run without forced oxidation of 2.2 lb/hr compares with 2.9 lb/hr for venturi/spray tower Run 907-1B with forced oxidation. Thus, it appears that air sparging for forced oxidation may increase adipic acid losses.

TEST RESULTS ON THE TCA WITH ADIPIC ACID ENHANCED LIME SCRUBBING

Although the majority of the adipic acid enhanced tests on the TCA were with limestone slurry, a single week-long test was conducted with lime slurry. Table 4 summarizes the results of this test along with base case lime tests without adipic acid.

Effect of Adipic Acid Concentration - The runs were conducted with lime on the TCA system, two without adipic acid and one with. Run 976-2A was a lime run at a scrubber inlet pH of 7 with 15 percent slurry solids concentration SO_2 removal averaged 70 percent. In Run 976-2B the slurry solids concentration was dropped to 8 percent and SO_2 removal dropped slightly to 67 percent. Finally, Run 977-2A was made with 8 percent slurry solids and nominally 300 ppm (actual average 420 ppm) adipic acid in the slurry liquor. SO_2 removal increased to 80 percent, an enhancement of 10 to 13 percentage points over the base cases with only a small addition of adipic acid.

It should be pointed out, however, that SO_2 removals in the mid 80's can be achieved in the Shawnee TCA system with lime alone by raising the scrubber inlet pH to 8 at 15 percent slurry solids. Additional tests will be conducted as time permits to work out the relationships between adipic acid concentration, scrubber pH, and slurry solids concentration.

Table 4

TCA SINGLE-LOOP TESTS WITHOUT FORCED OXIDATION
- ADIPIC ACID ENHANCED LIME SLURRY -

Major Test Conditions(1)	976-2A	976-2B	977-2A
Fly ash loading	High	High	High
Adipic acid concentration, ppm	0	0	300
Gas rate, acfm @ 300°F	30,000	30,000	30,000
Slurry rate, gpm	1200	1200	1200
Solids recirculated, wt.%	15	8	8
TCA inlet pH controlled at	7.0	7.0	7.0
EHT residence time, min.	4.1	4.1	4.1
EHT level, ft	17	17	17
Lime addition point(2)	DC	DC	DC
Run-Average Results			
Start-of-Run Date	7/12/78	7/19/78	7/25/78
Onstream hours	160	133	236
SO_2 removal, %	70	67	80
Inlet SO_2 concentration, ppm	2850	2950	2700
SO_2 make-per-pass, m-moles/liter	10.5	10.7	11.6
TCA inlet liquor gypsum saturation, %	90	85	90
Sulfite oxidation, %	16	12	10
Lime utilization, %	91	93	92
TCA inlet pH	7.0	7.0	6.95
Adipic acid concentration, ppm	0	0	420
Centrifuge solids, wt.%	63	63	61
Mist eliminator restriction, percent	0.2	1	1

(1) All runs were made with 3 beds and 5 inches per bed of 1-7/8 inch diameter, 11.5 gram nitrile foam spheres.
(2) DC = in the scrubber downcomer

Section 5

DEWATERING CHARACTERISTICS OF ADIPIC ACID ENHANCED LIMESTONE SLURRY AT SHAWNEE

Settling and dewatering characteristics of slurry solids are routinely monitored in the Shawnee laboratory by cylinder settling tests and vacuum funnel filtration tests. A comparison of the results of these monitoring tests for limestone slurry with and without adipic acid addition is presented in this section and summarized in Table 5.

Cylinder settling tests are performed with slurries containing 15 percent solids at room temperature in a 1000 ml cylinder containing a rake which rotates at 0.16 rpm. The initial settling rate and ultimate settled solids concentration are recorded as indices of dewatering characteristics. The initial settling rate is only a qualitative index of the solids settling properties. Design rates for sizing clarifiers must take into consideration the hindered settling rate as the solids concentrate. The ultimate settled solids from the cylinder tests represent the highest solids concentration achievable in a settling pond.

Funnel filtration tests are performed in a Buchner funnel with a Whatman 2 filter paper under a vacuum of 25 in. Hg. The funnel tests correlate well with the Shawnee rotary drum vacuum filter when not blinded but the funnel test cakes tend to have lower solids concentration.

Table 5 lists the effects of adipic acid on both oxidized and unoxidized limestone slurries. The data reported are for a range of adipic acid concentration of 1500 to 3000 ppm. All samples were with high fly ash loadings in which about 40 percent of the slurry solids was fly ash. All tests were conducted with samples containing 15 percent slurry solids.

As reported previously, settling and filtration characteristics of oxidized slurry are much superior to the characteristics of unoxidized slurry. The same trend exists with the slurry samples containing adipic acid.

The average initial settling rate for oxidized limestone slurry decreased from 1.0 cm/min with no adipic acid to 0.5 cm/min with adipic acid. However, this rate was still considerably higher than settling rates without forced oxidation. Without forced oxidation, the initial settling rate averaged 0.2 cm/min with or without adipic acid.

Adipic acid had little or no effect on ultimate settled solids or funnel test cake solids. The data indicated a slight decline in solids quality with adipic acid but the decline was small compared with the difference between limestone slurry with forced oxidation and without.

Table 5

COMPARISON OF SHAWNEE WASTE SLURRY DEWATERING CHARACTERISTICS WITH AND WITHOUT ADIPIC ACID ADDITION

Alkali	Fly Ash Loading	Oxidation	Adipic Acid	Initial Settling Rate, cm/min		Ultimate Settling Solids, wt %		Funnel Test Cake Solids, wt %	
				Avg.	Range	Avg.	Range	Avg.	Range
LS	High	Yes	No	1.0	0.6-1.5	72	62-86	72	65-88
LS	High	Yes	Yes	0.5	0.3-0.9	72	59-83	69	59-77
LS	High	No	No	0.2	0.1-0.5	54	41-67	57	48-66
LS	High	No	Yes	0.2	0.1-0.3	51	42-69	56	49-73

In addition to the funnel filtration tests, the filter cake solids from the rotary drum vacuum filter were monitored. During forced-oxidation operation with adipic acid enhanced limestone, the rotary drum filter cake solids concentration ranged from 80 to 87 weight percent solids (see Table 1). This range is no different than that obtained when operating with oxidized sludge in the absence of adipic acid.

In summary, the only significant effect of adipic acid addition on dewatering characteristics was a decline in initial settling rate with oxidized limestone slurry. These observations agree generally with those at the IERL-RTP pilot plant.

Section 6

PRELIMINARY ECONOMICS OF ADIPIC ACID ENHANCED SCRUBBING

At the request of the Shawnee Steering Committee, the Emission Control Development Projects group of TVA projected preliminary economics of adipic acid addition for forced-oxidation systems designed to achieve an average of 90 percent SO removal from high sulfur flue gas. The preliminary results indicate that both capital and operating costs are reduced by about 5 percent for a limestone system with 750 to 1500 ppm adipic acid compared with a limestone system with no additive.

Conditions for these preliminary evaluations were prepared by Bechtel and are presented in Table 6. Results are listed in Table 7. The evaluations were based on a 500 MW scrubbing facility incorporating forced oxidation and operating on flue gas from coal containing 4 wt.% sulfur. The evaluations included $5/dry ton for waste solids disposal but excluded land costs for a disposal site and costs for preparing land.

The cases evaluated were:

- Case 1 - A limestone base case operated at relatively high liquid-to-gas ratio and limestone stoichiometric ratio required to achieve 90 percent SO_2 removal. Operation at 90 percent SO_2 removal with limestone alone, although possible, has not been demonstrated at Shawnee. Two effluent hold tanks were included for forced oxidation in the first tank at lower pH before limestone is added in the second tank.

- Case 2 - A limestone case with MgO addition. Oxidation of the scrubber bleed stream was chosen in this case because oxidation within the scrubber loop is incompatible with magnesium enhanced scrubbing.

- Cases 3, 4, and 5 - Limestone cases with adipic acid addition. Adipic acid concentrations in the scrubber liquor of 750, 1000, and 1500 ppm were evaluated. Adipic acid consumptions of five times theoretical were used to allow for the level of unaccounted losses observed at Shawnee. Shawnee tests made after conditions were chosen for the economic evaluation indicate that SO removal for these cases should be higher than 90 percent. A single hold tank was chosen for these cases because the low scrubber inlet pH with adipic acid is compatible with forced oxidation.

Table 6

CONDITIONS FOR PRELIMINARY ECONOMIC ANALYSIS OF ADIPIC ACID ENHANCED LIME/LIMESTONE WET SCRUBBING WITH FORCED OXIDATION

Capacity:	500 MW
Coal:	4 wt % sulfur
Scrubber:	TCA with 3 beds, 4 grids, and 5 inchs sphere height per bed
Superficial Gas Velocity:	12.5 ft/sec
Scrubbing Mode:	Single loop with forced oxidation
Number of Scrubbing Trains:	4
Dewatering:	To 85 wt % solids by thickener and rotary drum vacuum filter
Sludge Disposal:	Includes $5/dry ton for waste solids disposal but excludes land costs for a disposal site and costs for preparing land
Onstream Factor:	7000 hours of operation/year

Case No.	1	2	3	4	5	6
Alkali	LS	LS	LS	LS	LS	Lime
Additive	-	MgO	Adipic Acid	Adipic Acid	Adipic Acid	Adipic Acid
Additive conc., ppm	-	5500[1]	750	1000	1500	1000
Additive rate, lb/hr	-	168	64[2]	85[2]	128[2]	80[2]
Percent SO_2 removal	90	90	90-95	90-95	95	95
L/G, gal/Mcf	58	50	50	50	50	50
Alkali stoic. ratio, mole Ca/mole SO_2 absorbed	1.55	1.20	1.20	1.20	1.20	1.05
Inlet pH	5.8	5.4	5.4	5.4	5.4	7.0
Filter cake solids, wt %	85	85	85	85	85	85
Sulfite oxidation, %	99	99	99	99	99	99
Air stoich., atoms O/ mole SO_2 absorbed	1.7	1.7	1.7	1.7	1.7	1.7
Mode of oxidation	in loop (2 EHT)	bleed stream	in loop (1 EHT)	in loop (1 EHT)	in loop (1 EHT)	in loop (2 EHT)

Notes: [1] Effective Mg^{++}
[2] Five times theoretical consumption

Table 7

RESULTS OF PRELIMINARY ECONOMIC ANALYSIS OF ADIPIC ACID ENHANCED LIME/LIMESTONE WET SCRUBBING WITH FORCED OXIDATION

Case	Alkali/Additive	Additive Conc., ppm	Average Percent SO_2 Removal	Total Capital Investment[3]		First Year Revenue Requirement[2]	
				$MM(1979)	$/kW	$MM(1980)	Mills/kWh
1	Limestone	-	90	41.5	83.1	20.9	5.96
2	LS/MgO	5500[1]	90	41.0	82.1	20.1	5.76
3	LS/Adipic Acid	750	90-95	39.3	78.5	19.6	5.60
4	LS/Adipic Acid	1000	90-95	39.4	78.7	19.8	5.64
5	LS/Adipic Acid	1500	95	39.5	79.0	19.9	5.69
6	Lime/Adipic Acid	1000	95	38.8	77.6	21.4	6.11

Notes: [1] Effective Mg^{++}

[2] Includes 17.2% annual capital charge

[3] Does not include land costs for a disposal site or costs for preparing land

Raw Materials Costs:
Limestone - $7/ton
Lime - $42/ton
MgO - $300/ton
Adipic Acid - $840/ton

- Case 6 - A lime case with adipic acid addition. As with the limestone cases, five times theoretical consumption of adipic acid was used to allow for unaccounted losses. At the conditions chosen, SO_2 removal may be as high as 95 percent. Two effluent hold tanks were included to allow forced oxidation in the first tank before lime addition in the second.

For the cases studied, as shown in Table 7, annual revenue requirements were lowest for the adipic acid enhanced limestone cases. The annual revenue requirements included a 17.2 percent annual capital charge.

Annual revenue requirement for limestone with 1500 ppm adipic acid (Case 5) was 5.69 mills/kWh compared with 5.96 for unenhanced limestone (Case 1), a savings of almost 5 percent.

Scrubbing economics were insensitive to adipic acid consumption. An increase of adipic acid concentration from 750 ppm (Case 3) to 1500 ppm (Case 5) increased annual revenue requirements by only 1.5 percent from 5.60 to 5.69 mills/kWh. These values include 5 times theoretical adipic acid consumption.

The annual revenue requirement for magnesium enhanced limestone (Case 2) was 5.76 mills/kWh, lower than unenhanced limestone. Capital charges for the magnesium enhanced limestone case were higher than for the adipic acid enhanced cases because of the additional equipment required for bleed stream oxidation relative to oxidation within the scrubber loop.

The annual revenue requirement for adipic acid enhanced lime (Case 6) was 6.11 mills/kWh, the highest of the cases evaluated. This high value reflects the high cost of lime relative to limestone and the additional hold tank required for forced oxidation.

It should be noted that the differences in annual revenue requirements among these cases are small. The principal conclusion from these preliminary evaluations is that adipic acid addition does not increase costs but, in fact, decreases them slightly. Furthermore, costs are insensitive to adipic acid consumption over the practical range expected in FGD scrubbing.

Section 7

SUMMARY OF CHARACTERISTICS OF ADIPIC ACID AS A SCRUBBER ADDITIVE

Based on testing at the IERL-RTP pilot plant and at the Shawnee Test Facility and on the preliminary economic evaluations conducted by TVA, the characteristics of adipic acid as a scrubber additive can be summarized as follows:

BENEFICIAL EFFECTS

- Significantly enhances SO_2 removal efficiency
- Increases alkali utilization, hence decreases waste solids disposal requirements
- When used with limestone, has projected lower capital and operating costs than unenhanced limestone or limestone/MgO
- Can be used with both lime and limestone in either conventional or forced-oxidation modes for both new and existing installations
- Is not adversely affected by chlorides as is the limestone/MgO process
- Does not significantly affect solids quality (filterability/settling rate) as can occur with high magnesium ion concentrations
- Should promote use of less expensive and less energy intensive limestone rather than lime
- With proper pH control, steady outlet SO_2 concentrations can be maintained even with wide fluctuations of inlet SO_2 concentrations

NEGATIVE ASPECTS

- Has unpleasant odor associated with adipic acid decomposition product
- Adipic acid decomposition requires adding up to 5 times that theoretically required (However, consumption over the ranges anticipated has negligible economic impact)
- Other possible secondary environmental effects have not yet been determined. Separate studies are underway to determine if any such problems might exist.

Section 8

FUTURE EPA TEST PROGRAM WITH ADIPIC ACID

The EPA/IERL-RTP test program with adipic acid enhanced scrubbing systems is ongoing. Figure 8 shows the full program outline. With contingencies it is anticipated that the program will be completed by the end of 1979 or early 1980. The following activities are either planned or proceeding:

- Tests at Shawnee to develop a single-loop forced oxidation system with adipic acid enhancement for both lime and limestone
- Factorial tests at Shawnee to develop the interrelationships of operating parameters for both lime and limestone and both with and without forced oxidation
- Tests at Shawnee to determine if the slurry bleed stream can be oxidized outside the scrubber loop
- Full scale demonstrations, both with and without forced oxidation, of adipic acid enhanced limestone scrubbing
- Studies by Aerospace Corporation of the handling and disposal characteristics of waste sludges produced at Shawnee with adipic acid enhanced scrubbing
- Level 1 bioassay studies by Litton Bionetics Corporation to determine biological activity, if any, from adipic acid addition
- Evaluations by TVA of the economics of adipic acid addition
- Studies by Radian Corporation to evaluate the unaccounted losses of adipic acid and to develop better analytical procedures
- A limited market study to determine the effect of extensive adipic acid consumption in FGD scrubbing on the adipic acid market.

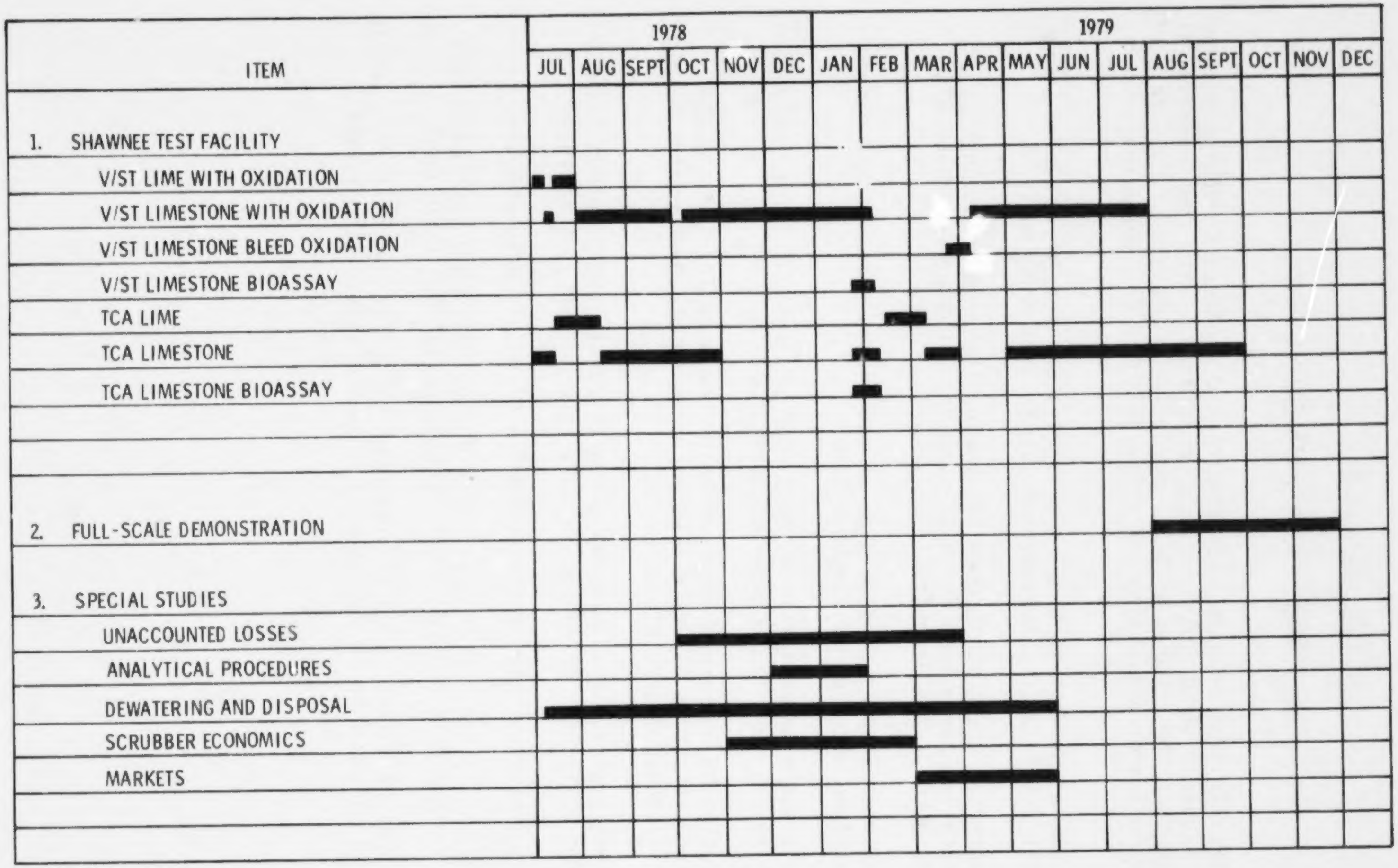


Figure 8. SHAWNEE ADVANCED PROGRAM PROJECTED TEST SCHEDULE

Section 9

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Appendix

CONVERTING UNITS OF MEASURE

Environmental Protection Agency policy is to express all measurements in Agency documents in metric units. In this report, however, to avoid undue costs or lack of clarity, English units are used throughout. Conversion factors from English to metric units are given below:

To Convert From	To	Multiply By
scfm (60°F)	nm/hr (0°C)	1.61
cfm	m/hr	1.70
°F	°C	(°F-32)/1.8
ft	m	0.305
ft/hr	m/hr	0.305
ft/sec	m/sec	0.305
ft^2	m^2	0.0929
ft^2/tons per day	m^2/metric tons per day	0.102
gal/mcf	$1/m^3$	0.134
gpm	1/min	3.79
gpm/ft^2	1/min/m^2	40.8
gr/scf	gm/m^3	2.29
in.	cm	2.54
in. H_2O	mm Hg	1.87
lb	gm	454
lb-moles	gm-moles	454
lb-moles/hr	gm-moles/min	7.56
lb-moles/hr ft^2	gm-moles/min/m^2	81.4
lb-moles/min	gm-moles/sec	7.56
psia	kilopascal	6.895

TVA COMPLIANCE PROGRAMS FOR SO_2 EMISSION

G. A. Hollinden
Energy Research
Tennessee Valley Authority
Chattanooga, Tennessee

C. L. Massey
Power Supply Planning Branch
Tennessee Valley Authority
Chattanooga, Tennessee

ABSTRACT

On April 19, 1976, a United States Supreme Court ruling favored mandatory constant SO_2 control requirements. Since that time, TVA has been implementing a billion dollar program to bring its power plants into compliance with State and Federal sulfur dioxide emission requirements. A consent decree was approved by the TVA Board of Directors on December 14, 1978, which will become effective upon approval by the Court. This paper will discuss the implementation of the compliance program at the TVA plants where flue gas desulfurization (FGD) will be used--Widows Creek, Paradise, Johnsonville, and Cumberland Steam Plants. Results associated with the Widows Creek unit 8 wet limestone scrubbing system will be presented. Coal washing, magnesium oxide scrubbing, and other innovative FGD processes will also be discussed.

TVA COMPLIANCE PROGRAMS FOR SO_2 EMISSION

TVA has either purchased lower sulfur coal or is installing needed control equipment to bring its coal-fired plants into compliance with sulfur dioxide emission requirements. The compliance plans for TVA's 12 coal-fired plants are based on the use of medium-or low-sulfur coal, the use of conventional coal-washing, and in the case of four plants, partial scrubbing of the plant flue gas. Four of these plants already meet emission standards, and compliance at three other plants will be achieved in 1979. At the five remaining plants, major installations of coal-washing facilities, baghouse collectors, or scrubbers are required; therefore, compliance will not be achieved until 1981 or later.

This paper will discuss compliance measures for those plants that require FGD systems--Widows Creek, Paradise, Johnsonville, and Cumberland.

WIDOWS CREEK STEAM PLANT

General Plant Description

TVA's Widows Creek Steam Plant is located on the Tennessee River in Jackson County, Alabama. The power plant consists of eight units (burning pulverized bituminous coal). These units have a combined generator capacity of about 1,977,985 kW. There are six units in one powerhouse; five of which have a capacity of 140,625 kW each, and one with a capacity of 149,850 kW. The other two units, one with 575,010 kW and a second with 550,000 kW output, are in a second powerhouse. The compliance schedule requires that by September 1981 the Widows Creek Steam Plant will meet a plantwide emission limitation of 1.2 lbs. SO_2 per million Btu's heat input on a 24-hour average basis. To meet this emission limitation, the six smaller units will burn a coal that ensures compliance with an emission limit of no more than 1.6 lbs of SO_2 per million Btu, and limestone scrubbers will be used on Units 7 and 8 that will reduce SO_2 emissions to no more than 0.9 lbs SO_2 per million Btu's for these units.

A full-scale wet limestone scrubbing system was installed on Unit 8. It was completed in 1977 and is now in full operation. The background history and design premises were discussed in a publication by McKinney, Little, and Hudson.[1] Operating experience was recorded in a publication by Wells, Muirhead, and Buckner.[2] The data in these publications will not be repeated, but results previously reported will be updated with emphasis on the operating problems associated with running this scrubbing system. Figure 1 is a flow diagram of the wet limestone scrubbing system.

SO_2 removal efficiences, which result in emission levels of 1.2 lbs or less per million Btu's, have been achieved. However, during the period that the scrubber has been in operation, the unit has operated at reduced load due to boiler problems. Several problems have prevented sustained operation of the scrubber system. The major problems and solutions are as follows:

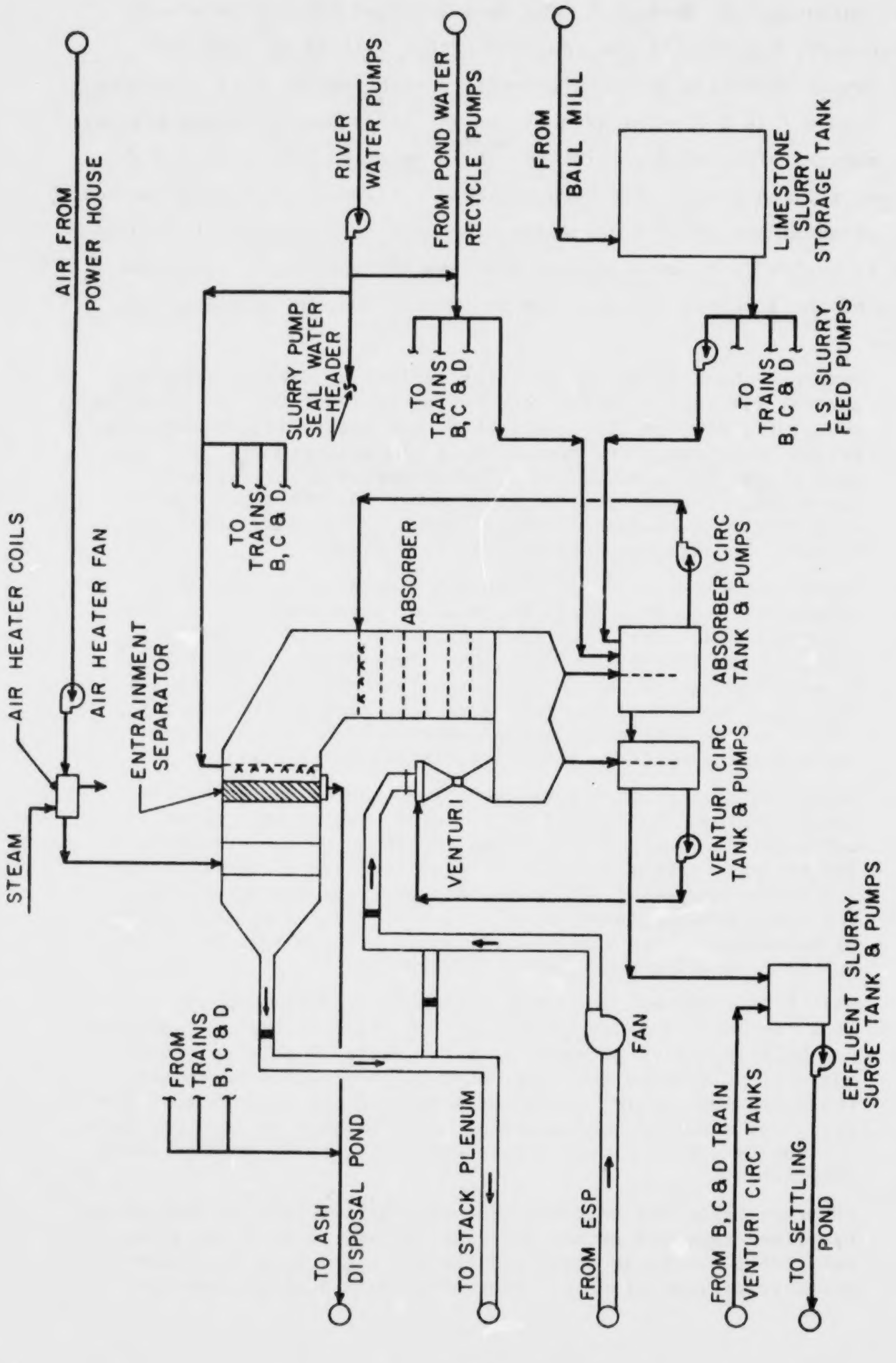
1. Dampers - Each scrubber has three guillotine dampers (one each for the gas inlet, outlet, and bypass ducts). Due to defective seals, fly ash and flue gas leakage has caused failure of electrical components, deterioration of protective boots, and jamming of the drive mechanism. Andco dampers were originally installed. Currently, a modified Andco and two other commercial dampers (Mosser and Damper Design Industries) are being tested. No data are available at this time.

2. Expansion Joints - The 56 expansion joints in the scrubber system are made of a 5-ply material of asbestos, teflon, and stainless-steel mesh. Solids, which have collected between the abrasion shield and joint, have caused the material to be worn through in several places with resulting gas leakage. Operating experience has indicated that some of the expansion joints can be eliminated while others will be modified by replacing the material with flexible stainless steel.

3. Rubber Linings - To prevent corrosion, the scrubbers have a rubber lining. Failure of this lining, due to poor metal adhesion, seriously affected the operation of the scrubbers. The failures occurred on the sloping sections of the absorber and venturi hoppers. To solve this problem, the linings in the sloping sections were replaced with 316-L stainless-steel plate welded to the carbon-steel shell. The remainder of the rubber lining has operated satisfactorily.

4. Ball Mill - The wet limestone ball mill was designed for 50 tons per hour, but during initial operation excessive limestone and ball rejection occurred at rates above 20 tons per hour. Inspection revealed that the metal helix designed to retain the balls and control limestone rejection was inadequate. The helix was modified, and grinding rates of over 40 tons per hour can now be achieved with a size consisting of 90 percent minus 200 mesh.

5. Absorber Grids and Nozzles - The absorber section of each train has five trays and grids. Originally, the three lower grids were 316-L stainless steel and the two upper ones were fiberglass reinforced plastic. Poor spray distribution from the


AIR FROM POWER HOUSE
RIVER WATER PUMPS
FROM POND WATER RECYCLE PUMPS
FROM BALL MILL
LIMESTONE SLURRY STORAGE TANK
AIR HEATER COILS
AIR HEATER FAN
SLURRY PUMP SEAL WATER HEADER
TO TRAINS B, C & D
TO TRAINS B, C & D
TO TRAINS B, C & D
LS SLURRY FEED PUMPS
ENTRAINMENT SEPARATOR
ABSORBER
ABSORBER CIRC TANK & PUMPS
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VENTURI
VENTURI CIRC TANK & PUMPS
FROM TRAINS B, C & D
TO ASH DISPOSAL POND
TO STACK PLENUM
FROM ESP
FAN
EFFLUENT SLURRY SURGE TANK & PUMPS
FROM B, C & D TRAIN VENTURI CIRC TANKS
TO SETTLING POND

Figure 1. Flow Diagram of Widows Creek Unit 8 Scrubber System

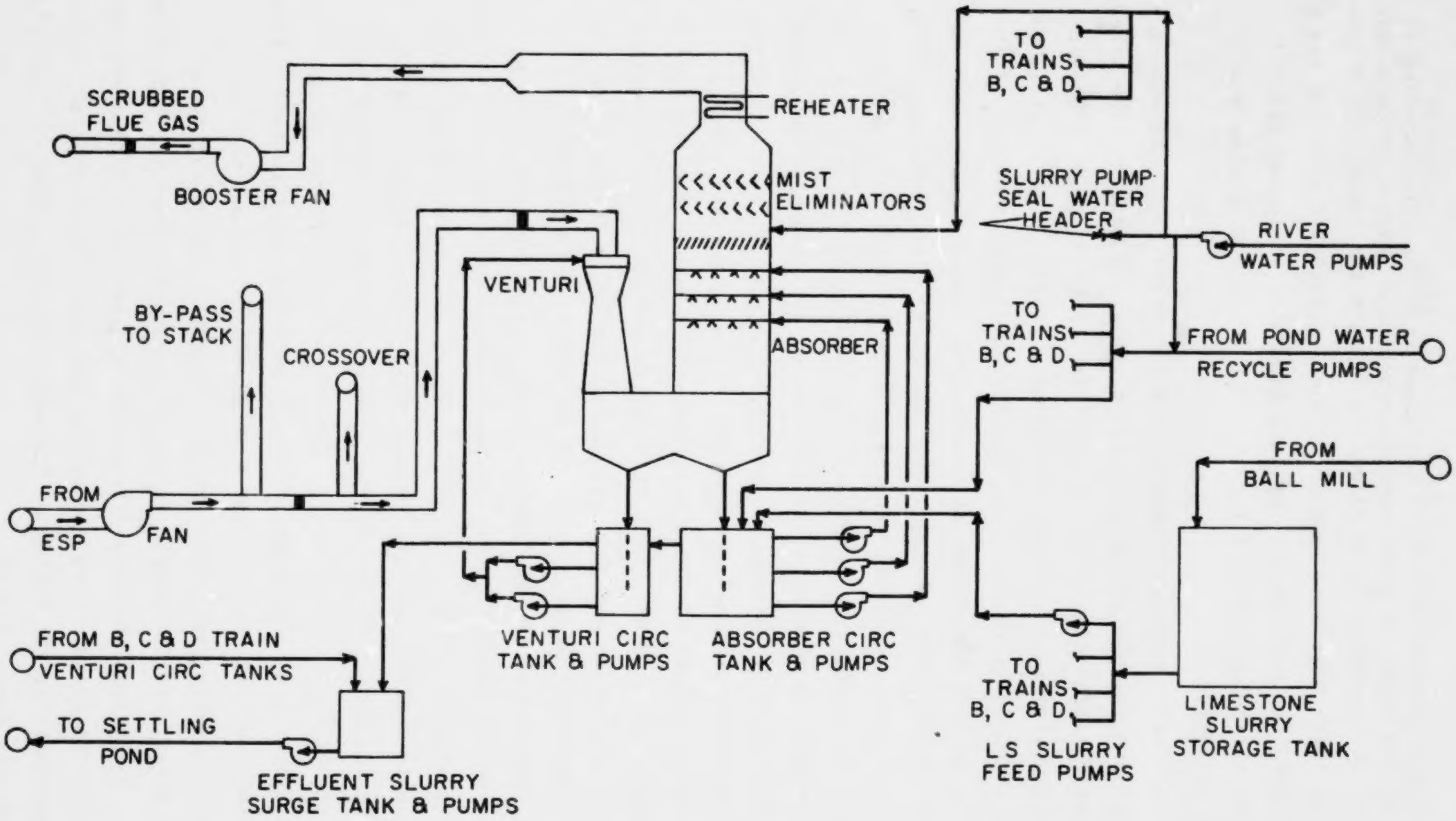
nozzles caused erosion of the plastic grids. The location of the two types of grids were reversed, and several spray nozzles are being tested in an effort to improve the slurry distribution.

Although there have been several problems associated with the scrubbing system, availability and operability have gradually increased since startup. The total capital cost for the Unit 8 scrubber system was \$54 million (\$100/kW).

A contract was awarded to Combustion Engineering, Inc., to supply a limestone FGD process for the 575,010 kW Widows Creek Unit 7. The schedule for this unit requires completion of installation by March 1, 1981, and shakedown and performance tests by September 1, 1981. A bonus will be paid if the schedule is improved and liquidated damages assessed if it is delayed. The cost of the system as provided by Combustion Engineering is \$23,555,200; however, TVA estimates that the total cost, including TVA's internal expenditures, will will be \$54 million or about \$100/kW.

Figure 2 is a schematic drawing of the FGD system for Unit 7. The system is designed to remove particulate matter and sulfur oxides. The flue gas, after passing through an electrostatic precipitator, is drawn into the particulate removal section, which is a converging section containing staggered layers of 317-L stainless-steel rods. The vertical spacing between the rods will be controlled to maintain a predetermined pressure drop to provide maximum particulate removal. This section, in addition to removing a large portion of the particulate matter, is responsible for some of the SO_2 removal. The scrubbing slurry in the particulate removal section is introduced by means of replaceable jar-type ceramic spray nozzles. A portion of the slurry is sprayed on the inlet walls to keep them clean, and the remainder is sprayed directly on the rods. To prevent solid buildup at the wet-dry interface, a steam soot blower is located in the inlet duct. The run off slurry from the particulate removal section is discharged directly into the reaction tank. From the particulate removal section, the flue gas turns, passes through ladder vanes to assure even distribution, and enters the spray tower.

In the spray tower, multiple stages of sprays are used to distribute the slurry. The ceramic nozzles atomize the slurry into a fine spray which provides the large surface area needed for the mass transfer of SO_2 into the liquid. Upon reaching the bottom of the absorber, the slurry is discharged directly into the reaction tank.



SCRUBBED
FLUE GAS
BOOSTER FAN
REHEATER
MIST
ELIMINATORS
TO
TRAINS
B, C & D
SLURRY PUMP
SEAL WATER
HEADER
RIVER
WATER PUMPS
VENTURI
BY-PASS
TO STACK
CROSSOVER
ABSORBER
TO
TRAINS
B, C & D
FROM POND WATER
RECYCLE PUMPS
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BALL MILL
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SURGE TANK & PUMPS
L S SLURRY
FEED PUMPS

Figure 2. Flow Diagram of Widows Creek Unit 7 Scrubber System

The cleaned flue gas passes through the first stage separation section called the bulk entrainment separator (BES). The BES consists of 6-inch fiberglass reinforced plastic vanes mounted at a 45-degree angle on 2-inch parallel spacing. The upper- or mist-eliminator section is constructed of Vee-shaped fiberglass reinforced plastic vanes arranged in a series of chevrons across the gas flow path. Two rows of chevrons are used to minimize mist carry over. Retractable water lances, which rotate 360 degrees with pairs of opposed nozzles at the ends and mid points and which are located between the BES and the lower-level chevrons, are used to clean the mist eliminators.

The temperature of the flue gas leaving the mist eliminators is about 120°F, and almost all the entrained moisture has been removed. However, the gas is saturated and contains a small amount of entrained moisture which, if condensation occurs, would have a detrimental effect on the induced draft fans, ductwork, and stacks. To prevent condensation and the resultant problems, the gas passes up through a finned-tube reheater which increases its temperature about 50°F. The gas travels from the reheaters through the outlet ductwork, to the ID fans, and then to the stack.

The reaction tanks, located below the particulate-removal section and the absorber, are constructed of carbon steel. Retention of the spray solutions in the reaction tanks allows time for the completion of chemical reactions and the precipitation of calcium sulfite and sulfate. The required make up water and limestone additive are piped to the absorber reaction tank. The solution is recirculated from the reaction tanks to the spray nozzles with rubber-lined spray pumps.

To convert the calcium sulfite to calcium sulfate, forced-oxidation equipment is being tested on one train of Unit 8. If successful it will be installed on both Units 7 and 8. The calcium sulfate which is formed when calcium sulfite is oxidized can be more easily disposed of than the sulfite.

A portion of the slurry is bled off from the particulate-removal reaction tank to provide necessary solids removal from the cycle. The bleed rate to the disposal system is regulated to maintain the proper slurry concentration.

Disposal of the solids produced by the FGD systems on Widows Creek Units 7 and 8 will originally be accomplished by ponding. The 110-acre pond presently serving Unit 8 is expected to provide adequate storage for disposal of the sludge from both Units 7 and 8 until late 1983. Studies on sludge minimization, further handling, and fixation will be conducted; and a decision about optimal long-range sludge disposal and land reclamation will be made before commercial operation of Unit 7. The alternatives to ponding raw sludge which are being investigated are dewatering, mixing, layering, oxidation, and fixation.

PARADISE STEAM PLANT

General Plant Description

TVA's Paradise Steam Plant is located in Muhlenburg County, Kentucky, on the Green River. The plant consists of three steam-electric generating units fueled by bituminous coal. The total plant maximum generator nameplate rating is 2,558,000 kW. Units 1 and 2, which began commercial operation in 1963, are each rated at 704,000 kW. These two units are equipped with 600-foot smokestacks and electrostatic precipitators which control particulate emissions to 0.20 lb fly ash/10^6 Btu. Unit 3, which is rated at 1,150,000 kW, began commercial operation in 1970. This unit has an 800-foot smokestack and an electrostatic precipitator which is presently limiting emissions to approximately 0.50 lb fly ash/10^6 Btu. In order to bring Unit 3 into compliance with Kentucky's particulate-emission standard of 0.11 lbs fly ash/10^6 Btu, a new precipitator is being added in series with the existing precipitator.

The Paradise Steam Plant is frequently called a "mine mouth" plant, because it is located in the midst of major coal fields in western Kentucky. The coal in this area is relatively low quality. In 1977 the average analysis was: Ash 17.3 percent, Sulfur 4.2 percent, and Heating Value 10,500 Btu/lb. The 4.2 percent sulfur is equivalent to 8.0 lbs SO_2/10^6 Btu. To comply with the proposed Kentucky emission standard of 3.1 lbs/10^6 Btu (3-hour average) for the plant, a coal-washing facility is being installed that will reduce the sulfur in coal to an extent that would limit sulfur dioxide emissions to 5.2 lbs/10^6 Btu. In addition, FGD systems will be installed on Units 1 and 2 that will further reduce SO_2 emissions from these units to no more than 0.9 lbs SO_2/10^6 Btu (3-hour average). Thus, the weighted average for the three units will achieve the SO_2 standard for the overall plant of 3.1 lbs SO_2/10^6 Btu (3-hour average). The scrubbers that will be installed on Units 1 and 2 will also have particulate-removal sections that will reduce the particulate emissions to the required 0.11 lbs/10^6 Btu.

Coal Cleaning Plant

John T. Boyd Company was commissioned by TVA to determine the washability of existing Paradise Coal as well as other midwestern coals most likely to supply the plant in the future. The Boyd Company submitted a report of its investigation which concluded the following:

1. The sulfur content of existing Paradise Coal could be reduced sufficiently by conventional washing methods to achieve a plant emission rate of 5.2 lbs $SO_2/10^6$ Btu.
2. The burning characteristics of the washed coal would be compatible with the requirements of the Paradise plant.
3. A single coal-washing plant at the Paradise site could be designed with sufficient flexibility to adequately process a variety of coals from the major midwestern coal seams.

Engineering design studies were made which indicate that it is feasible to locate a coal-washing facility at the Paradise plant and interface the plant with the coal-receiving facilities for the plant. The type of coal-washing facility will basically consist of a mechanical-dense medium-gravity separation process capable of removing large portions of the ash and sulfur from the coal (see Figure 3). The gravity separation process utilizes the fact that the mineral impurities (rock, slate, pyrites) existing in coal typically have higher specific gravities than the coal itself. In fact, the mineral pyrite (FeS_2), which is the source for a major portion of the total sulfur in typical west Kentucky coal, has a specific gravity four times that of pure coal.

Before the coal and its impurities can be separated, it must be run through a crusher which physically breaks the impurities away from the coal. This is possible because the seam joining the impurity with the coal is usually weaker than either of the adjoining materials. The result is a fracture at the seam. Once the coal is crushed and the impurities broken away, the mixture is introduced into a dense liquid medium usually consisting of finely ground magnetite suspended in water. The magnetite increases the apparent specific gravity of the water sufficiently enough to cause the coal particles to float; yet, it allows the more dense particles of impurities to sink. Magnetite is commonly used in the slurry because of its magnetic properties which ease its recovery for recirculation.

After the coal and refuse have been separated, the majority of the moisture will be drawn off by centrifuges. The refuse from the washing plant will be transported to a disposal area. Thermal drying of the coal may be required before conveying it to the steam plant for burning. The coal will be transported from the present plant storage and receiving system to the washing facility and then back to the plant on belt conveyors.

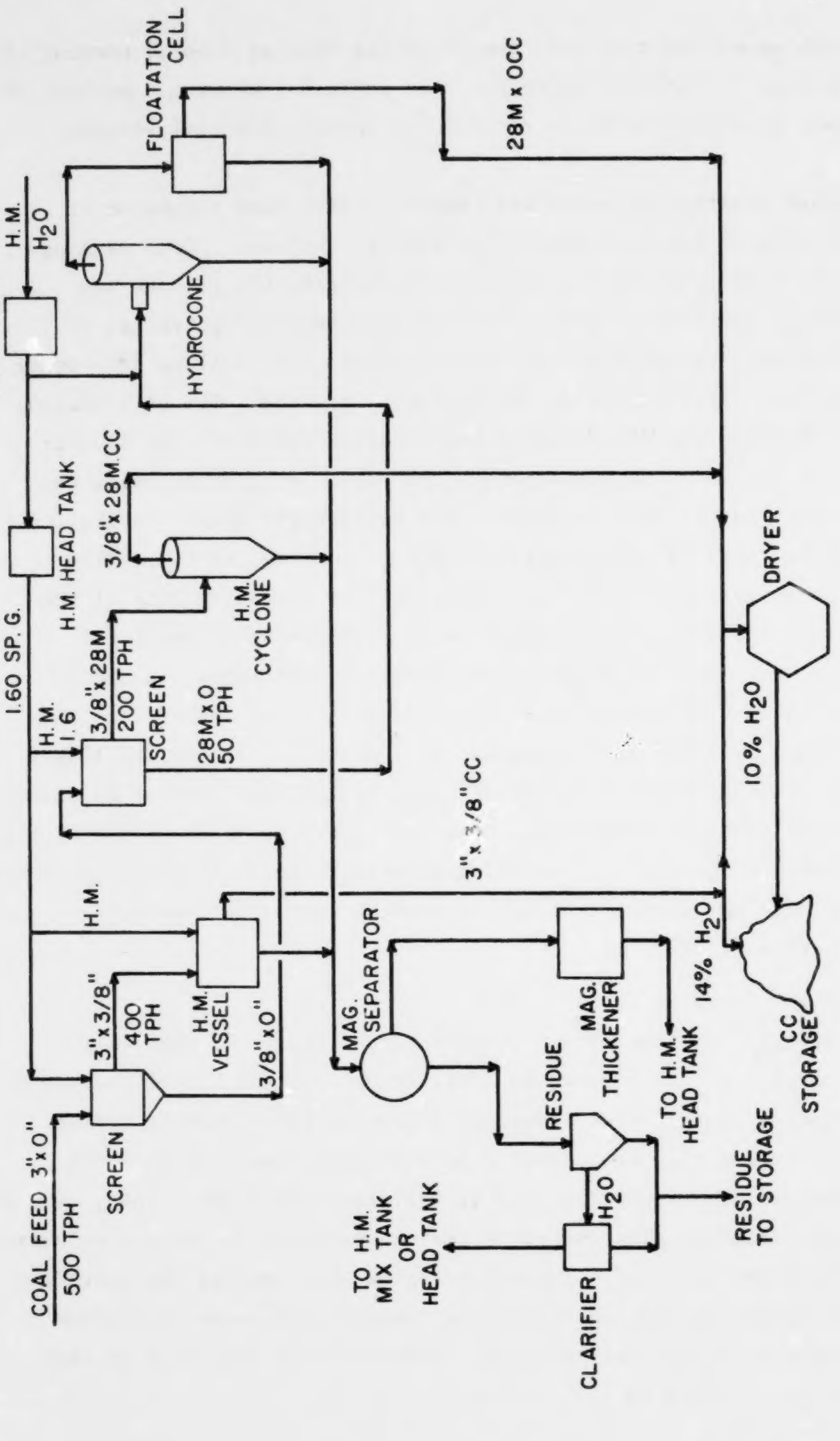
COAL FEED 3"x0"
500 TPH
SCREEN
3"x3/8"
400 TPH
3/8"x0"
H.M. VESSEL
H.M.
MAG. SEPARATOR
RESIDUE
MAG. THICKENER
TO H.M. HEAD TANK
H_2O
CLARIFIER
TO H.M. MIX TANK OR HEAD TANK
RESIDUE TO STORAGE
14% H_2O
CC STORAGE
3"x3/8"CC
H.M. 1.6
1.60 SP. G.
SCREEN
3/8"x28M
200 TPH
28Mx0
50 TPH
H.M. CYCLONE
H.M. HEAD TANK
3/8"x28M CC
MIX TANK
H.M.
H_2O
HYDROCONE
FLOATATION CELL
28M x OCC
DRYER
10% H_2O

Figure 3. Heavy Media Coal Cleaning System

The washing of the coal will result in the loss of 4 to 15 percent of the heat content of the coal as refuse. In addition, about 0.8 percent of the net plant generation will be required to operate the coal-washing facilities.

There are substantial potential benefits other than reduction of sulfur content that may be derived from washing the coal. Coal washing will significantly reduce the amount of ash entering the furnace and possibly change ash composition. These changes have the potential of improving reliability and availability of the plant as well as decreasing maintenance cost attributable to the poor coal quality. The coal-washing costs will be offset by the value of improvements realized. It is estimated that improvements in operation of the Paradise plant by using the clean coal will result in a savings of $16 million per year. Capital cost of the washing facility and related equipment is estimated to be $130 million. Amortization of this investment for the remaining life of the Paradise plant (32 years) will amount to an annual cost of about $17 million. In addition, the Boyd report estimated that operation and maintenance costs will amount to about $1.50 per ton of washed coal and that Btu losses will average approximately 7 percent. If the Btu losses are evaluated based on an estimated raw coal replacement cost of $1.00/10^6 Btu, the annual cost of washing Paradise coal would total about $32 million. If the savings of $16 million resulting from improvement of plant operation is realized, the net annual cost of the washing facilities would be reduced to $16 million.

FGD Systems for Paradise Units 1 and 2

The flue gas desulfurization equipment (see Figure 4) that will be installed on each of the 704,000 kW Paradise Units 1 and 2 will consist of venturi-absorber systems with limestone slurry as the scrubbing medium. Six scrubber trains will be installed on each unit, any five of which will accommodate all of the flue gas at full-load operation. Thus, 1408 mW of scrubbing capacity, plus two spare trains providing 20 percent redundant capacity, will be installed. The spare trains will improve the operating reliability of the system and reduce the frequency of excess emissions resulting from equipment malfunctions. Induced draft fans will be used to draw the gas through the FGD system.

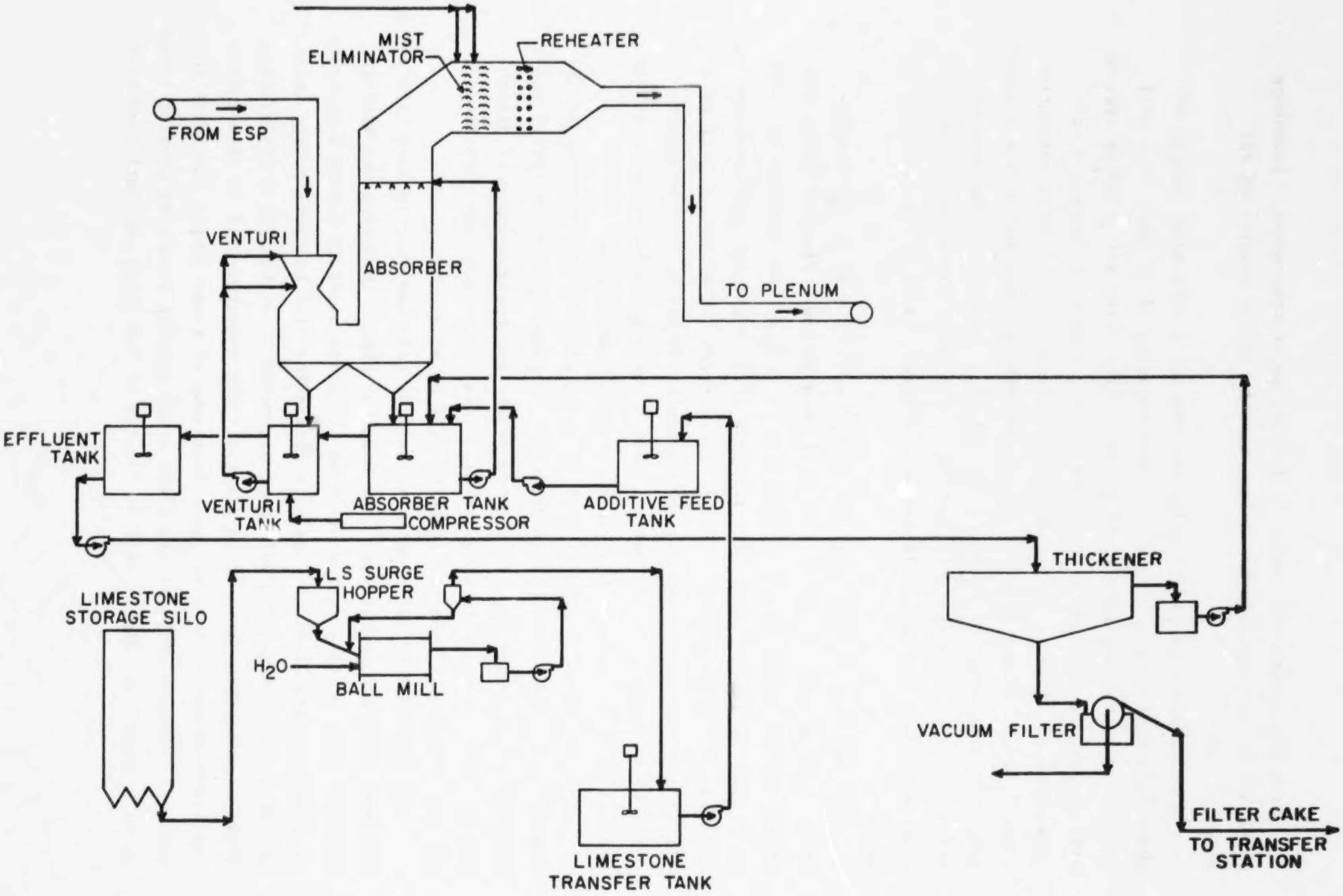
MIST ELIMINATOR
REHEATER
FROM ESP
VENTURI
ABSORBER
TO PLENUM
EFFLUENT TANK
VENTURI TANK
ABSORBER TANK
COMPRESSOR
ADDITIVE FEED TANK
THICKENER
L S SURGE HOPPER
LIMESTONE STORAGE SILO
H_2O
BALL MILL
VACUUM FILTER
LIMESTONE TRANSFER TANK
FILTER CAKE
TO TRANSFER STATION

Figure 4. Paradise FGD System

Each FGD system will consist of three major components: limestone preparation, venturi-absorber units, and the sludge dewatering and disposal train.

The limestone preparation for the two units will each consist of three ball mills, each having a minimum capacity of 35 tph. This will provide an excess capacity of 50 percent. The three wet grinding systems will operate in a closed circuit and produce a material having a size consisting of at least 90 percent minus 200 mesh and a slurry concentration of 40-60 percent solids. The slurry storage tank for each FGD system will hold enough for eight hours of full-load operation. The limestone slurry transfer from the storage tank to the venturi/absorber scrubber will use a loop for recirculation with control valves for each scrubber train.

The venturi/absorber will be located after the existing electrostatic precipitators; the ID fans will then follow. Plenums before and after the venturi/absorber will be used to control the turndown in order to follow the boiler load. The turndown will be accomplished by removing individual venturi/absorber trains from service. The venturi will have a slurry distribution system and an adjustable throat to ensure optimum particulate removal at all boiler loads. The high velocity areas of the venturi will be lined with abrasion-resistant materials.

The absorbers, in addition to spray nozzles for distribution of the slurry, will have chevron-type demisters to remove entrained slurry and a reheater to increase the temperature of the gas leaving the mist eliminator to 50°F. Washers to keep the mist eliminator clean and soot blowers to remove deposits from the reheater will be provided.

The material of all components will be selected to give long life and the most efficient operation. The scrubber shell, including the venturi section, will be constructed of 317-L stainless steel to ensure resistance to corrosion and erosion. The mist eliminator will be chevron-type units of two or more stages and will be constructed of either 317-L stainless steel or fiberglass reinforced plastic. The reheater will be bare-tube construction and will be at least four rows of tubes deep: the first four rows will be made of 317-L stainless steel and the remaining rows of carbon steel. Isolation dampers will be located at the inlet and outlet of each scrubber train to provide safe operating conditions under all operating modes for personnel working on equipment (fans, venturi/absorber, mist eliminator, reheater).

The sludge dewatering and disposal train will be designed to produce a filter cake $\geq$ 80 percent solids, without the addition of lime or other additive, for use as a physically stable landfill. To produce the 80-percent-solids filter cake, forced oxidation will be used in the scrubber system. Forced oxidation converts the calcium products formed in the reaction of SO_2 with limestone to gypsum. Gypsum can be dewatered to a minimum of 80 percent solids which can be used for landfilling. Forced oxidation will be accomplished by injecting compressed air into the scrubber reaction tanks through a manifold system of piping.

The dewatering system will consist of two thickeners, one for each unit, and three 50-percent-capacity vacuum filters for each thickener. The scrubber slurry will have a minimum of 8 percent solids and will be pumped to a thickener (which is the first step in the dewatering system). The solids settle in the thickener to a concentration of about 40 percent before being pumped to the vacuum filters, while the clarified liquor from the thickeners overflows into a trough and is returned to the scrubbers. The underflow from the thickeners is further dewatered in the vacuum filters to produce the 80-percent-solids filter cake.

The landfilling operation will consist of: transporting the dewatered cake to the disposal area; compacting, stacking, and covering the dewatered cake with soil as the disposal area is filled. The area will be revegetated with plant life compatible with the area and beneficial to wildlife. All aspects of the landfilling operation will be designed and operated to be in compliance with applicable regulations required by the Resource Conservation and Recovery Act and any other applicable regulations. Experimental evidence indicates that the use of oxidized-dewatered sludge will provide a method of overcoming environmental problems. It is estimated that approximately 200 acres of land will be needed to dispose of dewatered sludge for the remaining life of the Paradise Plant.

The capital cost of the FGD systems for the Paradise Units 1 and 2 is estimated to be $220 million. The annual cost of the project, including amortization of the capital over the remaining life of the plant and annual operating and maintenance costs, is estimated to be $62 million.

The schedule for installation of the two FGD units is as follows:

September 1, 1979	Initiate construction, Unit 1 FGD system.
December 1, 1979	Initiate construction, Unit 2 FGD system.
March 1, 1982	Complete construction and initiate shakedown operation on Unit 1 FGD system.
June 1, 1982	Complete construction of Unit 2 FGD system and complete shakedown and testing of Unit 1 FGD system.
September 1, 1982	Complete shakedown and testing of Unit 2 FGD system.

JOHNSONVILLE STEAM PLANT

General Plant Description

The Johnsonville Steam Plant, located in middle Tennessee, has 10 steam generators with a total capacity of 1,450,000 kW. Six of the generators, completed in 1953, have a capacity of 133,000 kW each and are tangentially fired with pulverized coal. The other four units, which started commercial operation in 1958, have a capacity of 162,000 kW and are rear-wall fired with pulverized coal.

The annual capacity factor of the total plant in 1977 was about 57 percent and is expected to decrease to 46 percent in 1983 and 25 percent in 1990, as more efficient plants come on stream. However, these overall annual factors do not show the peaking capability required by the Johsonville plant in the 1980's.

Fuel Supply

The primary fuel is bituminous coal from western Kentucky. The coal analysis range is 6.1-12.6 percent moisture, 13.5-20.5 percent ash, 1.8-4.4 percent sulfur, and a high heating value of 10,100-11,400 Btu per pound.

Compliance Requirement and Schedule

Johnsonville SO_2 emissions are to be limited to 3.4 lbs/10^6 Btu (3-hour average) by December 1, 1982.

The schedule to meet this requirement is as follows:

October 1, 1978	Award contracts for compliance coal (sulfur content equivalent to no more than 5.0 lbs SO_2 per million Btu).
March 1, 1979	Execute contract for FGD equipment.
September 1, 1979	Initiate onsite construction of FGD equipment.
September 1, 1982	Complete onsite construction of FGD equipment at 600 MW of capacity and begin final shakedown operation.
December 1, 1982	Complete shakedown operations on the 600 MW of FGD equipment. Achieve demonstrate compliance with the 3.4 lbs of SO_2 per million emission limit.

Compliance Program and FGD Description

It was decided that emissions at the Johnsonville plant would be limited to 3.4 lbs $SO_2/10^6$ Btu by using a combination of medium-sulfur coal and MgO scrubbers. The MgO scrubbers will be designed for a removal efficiency of 90 percent. Approximately 40 percent of the total plant flue gas will be scrubbed. For engineering reasons, a new 600-foot ground-based chimney will also be constructed.

A regenerative system was dictated because land was not readily available for disposal of the sludge from lime/limestone systems. Also, a portion of the sulfuric acid produced by a regenerative system could be used in the TVA fertilizer division. (No decision has been made on sulfuric acid marketing.) An MgO scrubbing system was chosen because the amount of commercial work with this system is greater than with any other regenerative system. (The plant arrangement is shown in Figure 5.)

The flue gas from the 10 steam generators will be fed into a plenum chamber from which 40 percent of the gas will be drawn through four scrubber modules (one module is a spare) where 90 percent of the SO_2 will be removed. The remainder of the flue gas which is unscrubbed and has a temperature of about 300°F will be mixed with the scrubber effluent gas before the total gas is fed to the single 600-foot stack. The concentration of SO_2 in the stack gas will be equivalent to no more than 3.4 lbs $SO_2/10^6$ Btu.

Description of Process. There are four scrubber modules, each of which is sized to scrub the flue gas from 200 MW of boiler capacity; one of the scrubber modules is a spare. Each scrubber module consists of two stages: one for chloride and fly-ash removal and one for absorption of SO_2. (A schematic diagram of the system is shown in Figure 6.)

The flue gas from the supply plenum is drawn into a venturi-type particulate scrubber where most of the fly ash and all the chlorides are removed. The gas is cooled from 300°F to the adiabatic saturation temperature (about 125°F). A portion of the particulate scrubber recirculating stream (a slurry of fly ash and mother liquor) is diverted to a pretreatment area, neutralized, and disposed of in the power plant ash disposal pond.

Flue gas enters the cocurrent spray tower after passing through a set of mist eliminators in the particulate scrubber. The flue gas is

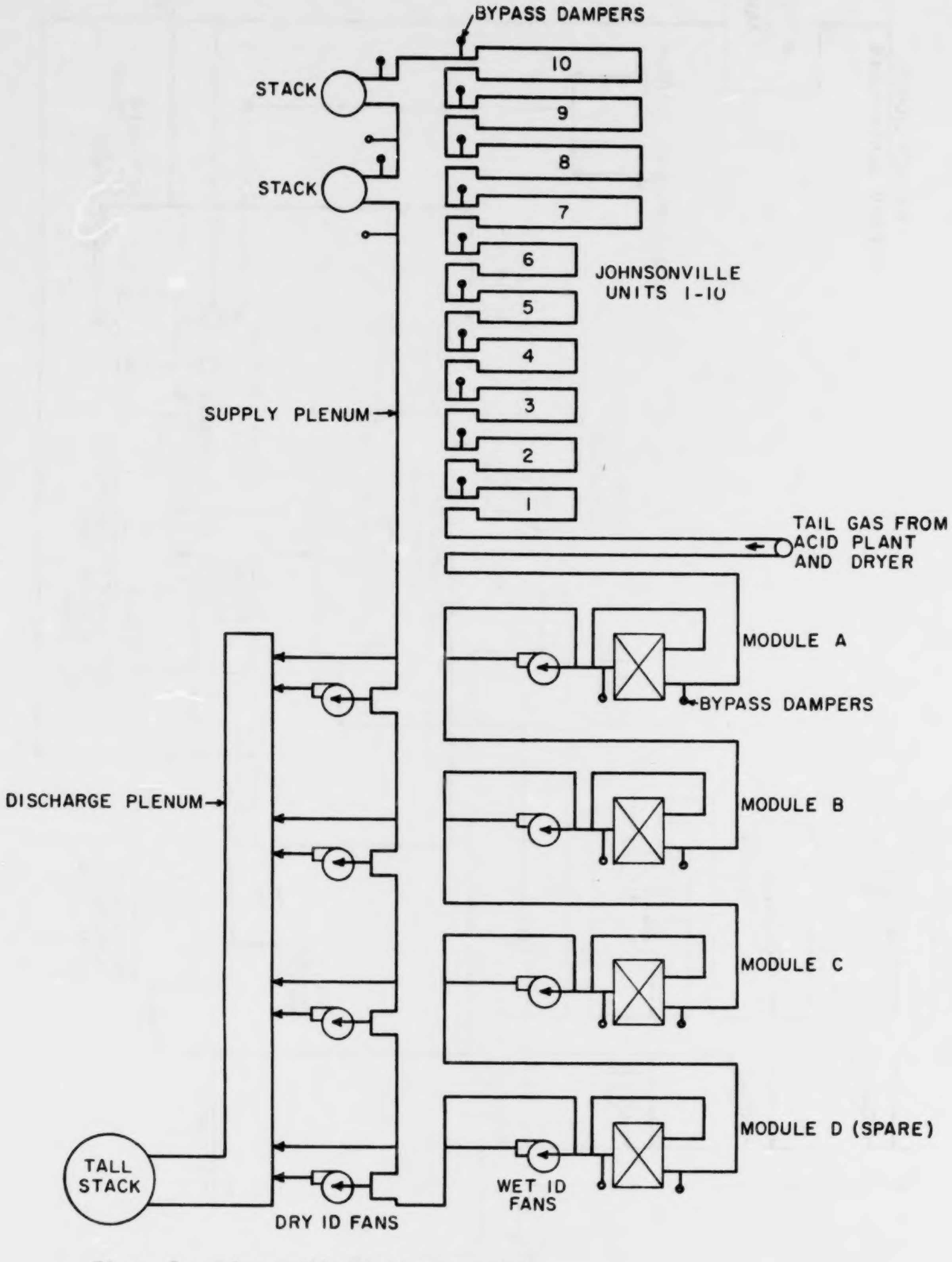
BYPASS DAMPERS
STACK
STACK
10
9
8
7
6
5
4
3
2
1
JOHNSONVILLE UNITS 1-10
SUPPLY PLENUM
TAIL GAS FROM ACID PLANT AND DRYER
MODULE A
BYPASS DAMPERS
DISCHARGE PLENUM
MODULE B
MODULE C
MODULE D (SPARE)
TALL STACK
WET ID FANS
DRY ID FANS

Figure 5. Johnsonville Plant Arrangement

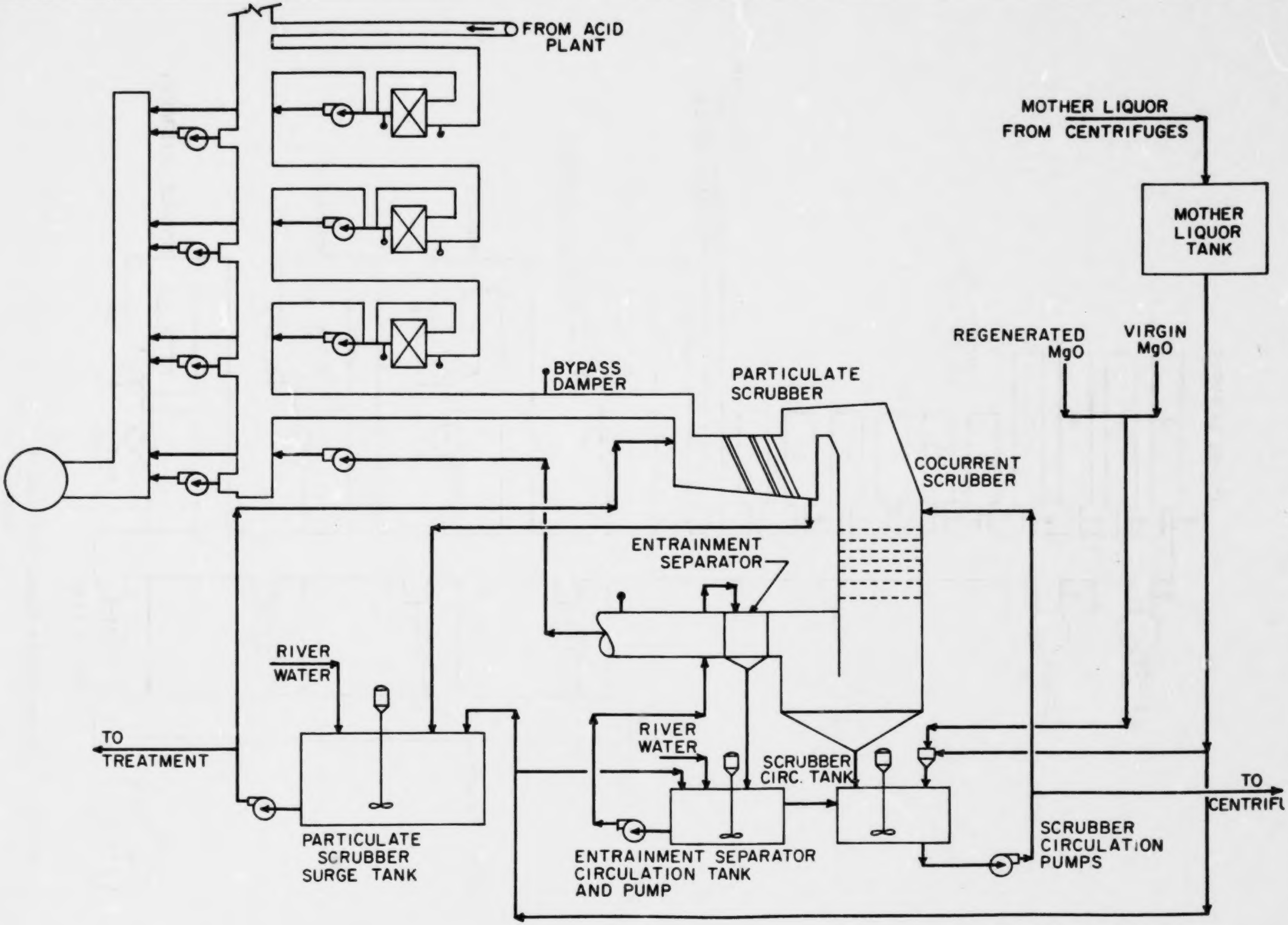
FROM ACID PLANT
MOTHER LIQUOR FROM CENTRIFUGES
MOTHER LIQUOR TANK
BYPASS DAMPER
PARTICULATE SCRUBBER
REGENERATED MgO
VIRGIN MgO
COCURRENT SCRUBBER
ENTRAINMENT SEPARATOR
RIVER WATER
TO TREATMENT
RIVER WATER
SCRUBBER CIRC. TANK
TO CENTRIFU
PARTICULATE SCRUBBER SURGE TANK
ENTRAINMENT SEPARATOR CIRCULATION TANK AND PUMP
SCRUBBER CIRCULATION PUMPS

Figure 6. Schematic Diagram of Johnsonville MgO Scrubbers

contacted with a recycling slurry of $MgSO_3$, $Mg(HSO_3)_2$, and $MgSO_4$ for the absorption of SO_2. The slurry contains the hydrated crystals of $MgSO_3$ and $MgSO_4$ as well as a solution saturated with each of these components. A purge stream off the recycled slurry containing approximately 10 percent solids is diverted to the centrifuges for separation of the solids from the mother liquor. The centrate is collected in the mother liquor tank and returned to the absorber and particulate-scrubber recirculating loops.

Each absorber is designed for a gas velocity of 15 feet per second, and a liquid-to-gas ratio (L/G) of 30 gallons per 1000 ft^3 of gas. Other design features include the following:

1. "Wet-elbow" design for removing a portion of the entrained slurry in the gas
2. Six (6) grids
3. Mixing of unscrubbed bypass flue gas with clean gas in the discharge plenum for reheat

The centrifuge cake is dried to remove waters of hydration in an oil-fired cocurrent rotary-kiln dryer. The dried ($MgSO_3$/$MgSO_4$) crystals are crushed, then transferred to an inprocess storage silo and before being fed to the fluid-bed calciner, the entrained solids in the dryer off gas are enclosed in a cyclone dust collector. The cleaned off gas, containing some SO_2 from partial breakdown of the $MgSO_3$, is returned to the scrubber area. (The MgO regeneration system is shown in Figure 7.)

The $MgSO_3$/$MgSO_4$ solids are weight fed to an oil-fired fluid-bed calciner. There they are calcined into MgO and SO_2. After discharge from the calciner, the SO_2-rich off gas passes through a cyclone dust collector where approximately 75-80 percent of the MgO dust is removed. The gas is then cooled from 1700°F to about 450°F by a series of heat exchangers. Next the cooled gas enters a baghouse, where the remaining MgO is removed. The SO_2 stream from the baghouse forms the feedstock for a conventional contact sulfuric acid plant. Tail gas from the acid plant is recycled to the SO_2 absorbers. All of the removed MgO is stored before recycling to the absorption area.

The principal objective in selecting the size of the acid plant and the storage capacity for $MgSO_3$ is minimal risk of restricting power plant operations. A double- or single-acid plant module with a total production capability of 350 tons per day is being investigated. Facilities

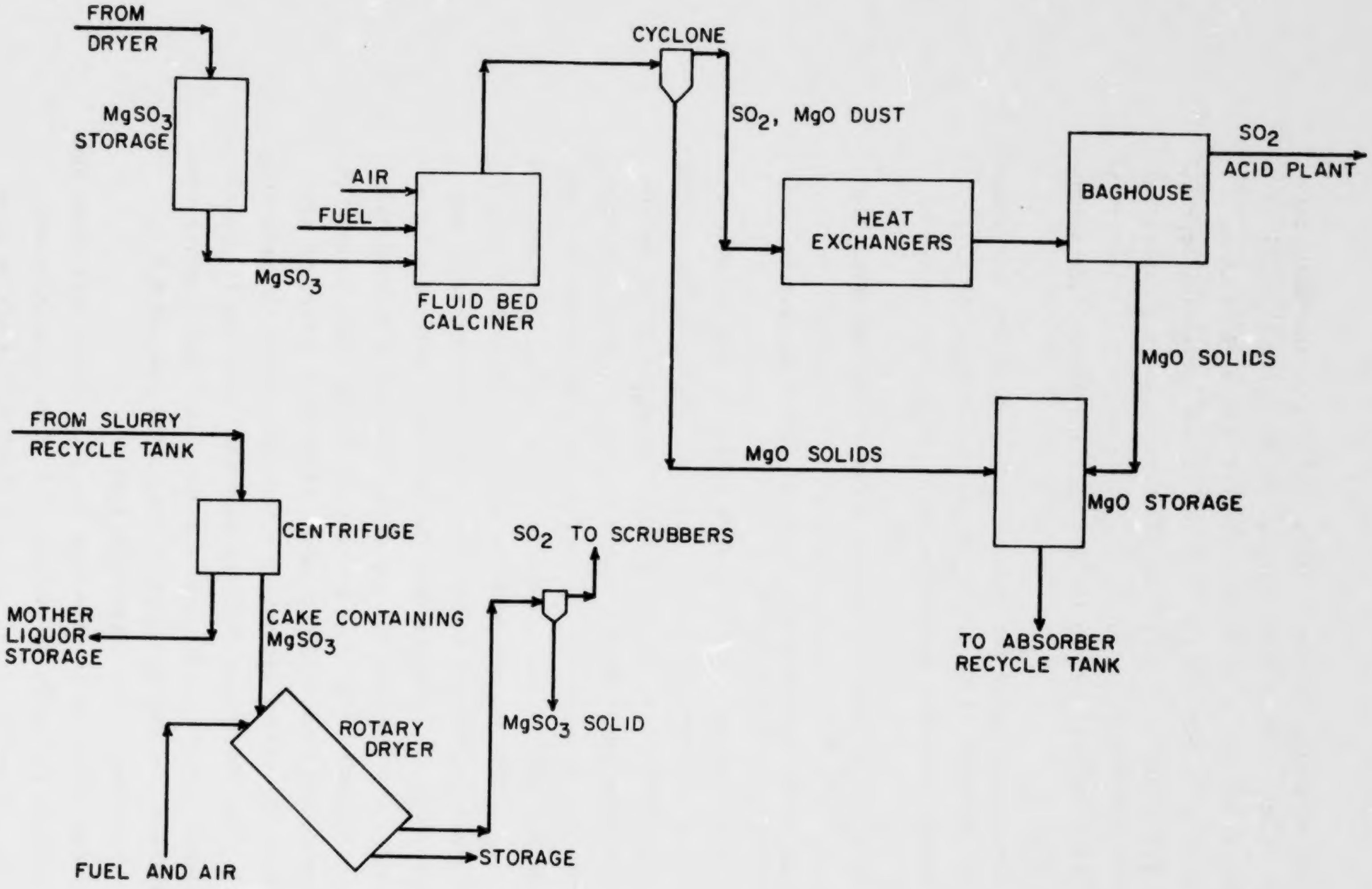
FROM DRYER
$MgSO_3$ STORAGE
AIR
FUEL
$MgSO_3$
FLUID BED CALCINER
CYCLONE
SO_2, MgO DUST
HEAT EXCHANGERS
BAGHOUSE
SO_2
ACID PLANT
MgO SOLIDS
MgO SOLIDS
MgO STORAGE
TO ABSORBER RECYCLE TANK
FROM SLURRY RECYCLE TANK
CENTRIFUGE
MOTHER LIQUOR STORAGE
CAKE CONTAINING $MgSO_3$
SO_2 TO SCRUBBERS
ROTARY DRYER
$MgSO_3$ SOLID
FUEL AND AIR
STORAGE

Figure 7. MgO Regeneration System

for storing $MgSO_3$ will be provided. Excess $MgSO_3$ storage equivalent to several-days' power plant operation appears quite adequate for minimizing the risk of restricting power plant operation. At this date, no decision concerning quantity has been made.

Estimated capital cost in 1982 dollars of an MgO scrubber system is $185,000,000. The cost includes the following:

1. Sulfuric acid plant and acid handling facilities
2. $MgSO_3$ and MgO storage
3. Five (5) scrubber modules
4. 600-foot stack
5. Drying and regeneration section
6. Ductwork tie-in for all ten units and scrubber ductwork
7. Site preparation*

The total annual cost including amortization is $36,000,000.

The annual operation and maintenance costs of the FGD system and acid plant are estimated to be $23,000,000. If it is assumed that $24 per ton will be received for the acid and about 256 tons per day will be produced, the operating and maintenance costs will then be reduced $2,240,000 to a net value of $20,760,000 (all 1982 dollars).

The estimated operation costs include a labor requirement of eight men per shift for the FGD system and acid plant. The 8-man shift does not include the maintenance manpower required.

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*When this paper was being written, TVA tentatively decided on a 4-module system. Although the technical changes have been incorporated in the paper, revised cost estimates are not available.

CUMBERLAND STEAM PLANT

The Cumberland Steam Plant is located approximately three miles northwest of Cumberland City, Tennessee, on the Cumberland River. It has two bituminous coal-fired steam generators, each with a capacity of 1,300,000 kW. The plant was initially placed into commercial operation in 1973 and is TVA's newest and largest coal-fired generating facility. The flue gas from the Cumberland plant is dispersed through two 1000-foot concrete stacks.

TVA is installing a heavy-media coal-cleaning system (such as the one shown in Figure 3) which will be capable of reducing SO_2 emissions from Cumberland to about 5.0 lbs/10^6 Btu. In addition, in a proposed settlement of litigation involving the Environmental Protection Agency, the States of Alabama and Kentucky, and a number of private citizen's groups, TVA has agreed to install at Cumberland a 600-MW FGD system with 90 percent removal (or the equivalent) of SO_2 to reduce plant emissions to 4.2 lbs/10^6 Btu. The proposed settlement provides that the deadline for installing these scrubbers may be extended by the parties if other than conventional lime/limestone scrubbers are installed.

To further develop advanced scrubber technologies and based upon TVA's ongoing program for the development and demonstration of these technologies, TVA is considering the feasibility of installing an alternative technology for the 600 MW's at Cumberland.

The selection of the system to install at Cumberland includes the evaluation of data obtained by testing of advanced scrubber concepts at three prototype levels (1 MW, 10 MW, and 20 MW) to evaluate process operation, performance of equipment and materials, and to obtain the optimum operating conditions for the processes. The objective of these tests and evaluations is to provide the most promising advanced scrubber system suitable for full-scale demonstration on the Cumberland plant. The following processes are being evaluated:

	Process	Absorbent
1.	Cocurrent Scrubber	Any
2.	Dowa-Double Alkali	Basic Aluminum Sulfate
3.	Thoroughbred 121-Gypsum	Sulfuric Acid-Limestone
4.	Absorption-Steam Stripping/Resox	Citrate
5.	Aqueous Carbonate	Sodium Carbonate

Primarily, these systems were chosen for evaluation because they do not produce a disposal byproduct such as the sludge from lime or limestone scrubbers. It should be noted that TVA has a comprehensive study now underway to develop the most environmentally acceptable method for disposal of the sludges produced by the limestone scrubbers now in operation at the Widows Creek plant and those to be installed at the Paradise plant. However, it is thought that the production of a useful byproduct is more desirable than the disposal product.

In addition to producing a useful byproduct, the five systems being studied have the potential for achieving high SO_2 removal efficiencies when compared to the limestone systems. When more efficient scrubber systems are developed and incorporated by EPA in its New Source Performance Standards, the resulting reduction in emissions will allow more room for growth of other sources.

In summary, we believe that the following alternatives offer not only a sound approach for reducing emissions at Cumberland but also for developing technology that will benefit the national interest as well as the TVA system.

Cocurrent Scrubber

In the cocurrent configuration, flue gas enters the scrubber from the top of the tower, where the liquid absorbent is also sprayed, and both flow cocurrently to the base of the scrubber. A majority of the liquid is removed from the gas by a wet-elbow scrubber bottom which forces the gas to make a turn of 180° after impacting the entrained liquid onto the surface of a pool of liquid maintained at the base of the scrubber. Thereafter, the gas passes through a chevron-type mist eliminator located in a horizontal duct with a separate water wash cycle. The gas is then reheated and discharged to the stack.

The cocurrent scrubber has several advantages over the conventional countercurrent scrubbers. Because of the significantly higher gas velocities permitted, the actual size of the scrubber will be smaller. The equipment configuration is more compatible with most power-plant duct and fan arrangements. The mist eliminator and reheater can be located near ground level and thus will be more easily maintained. Other potential advantages are better liquid and gas distribution and higher SO_2 removal efficiencies.

Tests of the cocurrent scrubber have been conducted at both the 1-MW and 10-MW levels with excellent results. The concept was found to be very flexible--capable of using almost any absorbent and very efficient (> 90 percent SO_2 removal) over a wide range of gas velocities (18-27 feet per second). The cocurrent scrubber is planned for the Johnsonville MgO system at a gas velocity of 15 ft/sec. If chosen for Cumberland, the cocurrent scrubber will most probably be designed for a gas velocity near the upper limit of the range described above.

Dowa Double-Alkali

In the Dowa process, SO_2 is absorbed in a clear solution of basic aluminum sulfate. The basic solution is oxidized by air in a separate tower and then neutralized with finely ground limestone to precipitate gypsum and to regenerate the basic aluminum sulfate solution. The process has been developed and tested on a 40-MW oil-fired boiler in Japan. However, it has not been tested on a coal-fired boiler.

Advantages of the Dowa process over the conventional lime/limestone scrubber are higher SO_2 removals and reduced scaling and plugging, possible because a solution rather than a slurry is used for scrubbing. The Dowa system produces a gypsum which can be stockpiled, used for landfill, or sold for making wallboard.

TVA plans to test the Dowa process at the 10-MW level with flue gas from a coal-fired unit. The prototype scrubber will be installed at the TVA Shawnee Test Facility. The project will be jointly funded by TVA, EPRI, and Universal Oil Products (licensee of the Dowa technology).

Chiyoda Thoroughbred 121 Process

In the Thoroughbred 121 process, the flue gas is quenched with water. It is then introduced into Chiyoda's patented Jet Bubbling Reactor where the flue gas is sparged into the absorbent through an array of vertical spargers, generating a jet bubbling (froth) layer. SO_2 is absorbed in the jet bubbling layer and subsequently oxidized to sulfate. The cleaned flue gas is discharged through a mist eliminator and out the stack.

Limestone slurry is pumped directly to the jet bubbling reactor where reaction takes place to produce gypsum. The same inherent advantages of gypsum production as discussed previously for Dowa are realized.

Tests of the Thoroughbred 121 are being conducted on a 20-MW coal-fired unit at Gulf Power Company's Scholz Steam Plant, Sneads, Florida, to evaluate performance, reliability, operability, and the cost and energy effectiveness of the process. This demonstration is a joint effort of EPRI, Southern Company Services, and Chiyoda. Testing began in August 1978. No reports have been published to date, but verbal reports from those involved indicate that excellent results are being obtained.

TVA will continue to follow the testing on the 20-MW plant at the Scholz steam plant. If the results continue to be favorable, TVA may propose to make a demonstration run under conditions more applicable to the Cumberland Steam Plant before deciding whether or not to proceed with a 600-MW unit at Cumberland.

Atomic International Aqueous Carbonate

In the Atomic International Aqueous Carbonate process, SO_2 in the flue gas is absorbed by sodium carbonate in a spray drier. The dried solids are separated from the flue gas in either an ESP or baghouse. The spent-absorber solids are discharged to a molten salt reactor where reduction of sulfite is accomplished. The "green liquor" material is then filtrated to remove insolubles and acidified in the carbonation step to produce sodium carbonate and H_2S. The sodium carbonate is returned to the scrubber and the H_2S is sent to a Claus plant for the production of elemental sulfur.

The advantages of this system are: (1) relatively high SO_2 removal; (2) spray drying requires no reheat; and (3) production of elemental sulfur.

If this alternative is selected, TVA will enter into a contract with Atomics International for supplying a 100-MW system at the Cumberland plant. Consequently, it would be necessary to use this system in combination with other FGD processes to reduce Cumberland emissions to the 4.2-pound level. The 100-MW size is necessary because of the advanced nature of the process, and the lack of opportunity for testing on a small scale makes it imprudent to commit to the full 600-MW level.

Absorption Steam Stripping/Resox Process

This process is a regenerable FGD system capable of producing elemental sulfur without a reducing gas or alternate byproduct. The process combines aqueous absorption with steam stripping.

The flue gas first goes to a venturi for particulate removal. From the venturi section it enters the bottom of the absorber and flows upward where it is contacted with a countercurrent flow of sodium citrate liquor to remove SO_2. After leaving the absorption section and mist eliminator, the gas is reheated and discharged to the stack.

The liquor containing the SO_2 is then steam stripped to remove the SO_2 and to regenerate the citrate liquor for reuse in the absorber. The SO_2 is then sent to the RESOX reactor where the SO_2 reacts with coal to form elemental sulfur.

The RESOX system is being developed by EPRI on a 42-MW scale at the Killerman Power Station of Steag A. G. in Lunen, Federal Republic of Germany.

The advantages of this process are: (1) high SO_2 removal; (2) the absorbent is reused; and (3) elemental sulfur is produced.

EPRI, TVA, and Flakt, Inc., are cooperating in testing the absorption steam-stripping process at TVA's 1-MW Colbert pilot plant, which will include short-term (factorial) and longer term operational testing to quantify all key process parameters. However, testing of this system may not be underway by the time a decision must be made for Cumberland. Therefore, evaluation of this system must be made on data from smaller scale studies. This again is an advanced process with little or no data except that obtained on the 1-MW test unit; therefore, it is thought that it would be imprudent to commit to a 600-MW unit.

The schedule for installation of the FGD system for Cumberland is dependent upon the route chosen for the type of system. By March 1, 1979, TVA must submit a notification of the type of FGD equipment (conventional lime/limestone or alternative) selected. If the conventional route is chosen, an invitation to bid must be extended by April 1, 1979, with award of contract by October 1, 1979. Onsite construction will be completed by October 1, 1982, and shakedown will be completed by December 31, 1982. If the alternative route is chosen, the FGD system to be installed will be selected by November 1, 1979, and an appropriate construction schedule developed.

SUMMARY

TVA has developed a program that will bring its coal-fired plants into compliance with State and Federal sulfur dioxide emission requirements. Eight of the twelve plants will meet SO_2 emission limits by selection of medium- and low-sulfur coals, but four plants will require coal washing and/or FGD systems to meet SO_2 standards. The plans for the four plants may be summarized as follows:

Plant and Size	Emission Limit	Compliance Program
Widows Creek Plant 8 units, 1978 MW	1.2 lbs SO_2 per 10^6 Btu (24 hour average)	a) Burn low-S coal on 6 units with total of 853 MW. b) Install limestone scrubber on 2 units totaling 1125 MW.
Paradise 3 units, 2558 MW	3.1 lbs SO_2 per 10^6 Btu (3 hour average)	a) Install coal-cleaning plant. b) Install limestone scrubber on Units 1 and 2, 704 MW each.
Johnsonville 10 units, 1450 MW	3.4 lbs SO_2 per 10^6 Btu (3 hour average)	a) Manifold all 10 units to a single stack. b) Burn medium-sulfur coal. c) Install 600 MW of MgO scrubbers and convert SO_2 to sulfuric acid.
Cumberland 2 units, 1300 MW	4.2 lbs SO_2 per 10^6 Btu (24 hour average	a) Burn washed coal with sulfur equivalent to 5.0 lbs $SO_2/10^6$ Btu. b) Install 600 MW of alternative scrubbers with 90% or more SO_2 removal efficiency or its equivalent (see text).

	Start Construction	Initial Operation[a]	Commercial Operation[b]	Compliance Dates[c]	
				Initial Operation	Compliance Test
Johnsonville					
MgO scrubber units 1-10	9/79	12/82	3/83	12/82	3/83
stack (600 ft) units 1-10	9/07/78[g]	NA	-	-	-
Widows Creek					
precipitators units 1-6	12/10/75	12/24/77	2/15/78[d]	3/01/78	-
limestone scrubber unit 8	2/20/73	4/30/77[e]	-	5/01/77	-
limestone scrubber unit 7 (contract)	9/01/78	2/81	5/81	3/81	9/81
Paradise					
coal washing facilities (contract)	6/01/77[f]	12/01/80	6/01/81	12/80	6/81
limestone scrubber units 1-2 (contract)	7/79	4/82	7/82	4/82	7/82
precipitator unit 3 (contract)	6/01/78[h]	6/80	9/80	9/80	11/80
Cumberland					
precipitators units 1-2	3/79	12/81	2/82	12/81	2/82
scrubbers[i]	7/01/79	7/01/82	12/01/82	7/82	12/31/82

[a] Initial operation or energization of last unit in project. (This date generally corresponds to the "Comp. Const" date on current TVA key date schedules.)

[b] Completion of startup testing of last unit in project. (This date generally corresponds to the "Comp Test" date on current TVA key date schedules.)

[c] Compliance with EPA standards.

[d] Completed demonstration test.

[e] First gas was scrubbed 5/16/77; slurry circulation through train "A" began 4/30/77.

[f] TVA site preparation work began 6/77; PA was approved 10/27/77; Contractor reported to the site 1/5/78 and began construction 3/10/78.

[g] TVA started conduit relocation 9/7/78; TVA to start foundation preparation 3/79; Contractor to start work 7/79.

[h] Started parking lot work 6/1/78; Contractor started work 7/10/78.

[i] This schedule is for the conventional lime/limestone scrubbing option.

TABLE 1. SCHEDULE OF COMPLIANCE METHODS

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SO_2 AND NOx REMOVAL TECHNOLOGY IN JAPAN

Jumpei Ando
Faculty of Science and Engineering
Chuo University
Kasuga, Bunkyo-ku, Tokyo 112

ABSTRACT

The total operational FGD capacity in Japan has reached 33,000 MW equivalent with about 50 plants for utility boilers and numerous smaller ones for other flue gas sources. Major FGD plants have been operated with over 97% operability removing over 90% of SO_2. Through FGD, hydrodesulfurization of heavy oil, and import of low-sulfur fuels, ambient SO_2 concentrations have been reduced remarkably since 1967 -- to levels low enough to meet the stringent air quality standard of 0.04 ppm in daily average or 0.02 ppm in yearly average.

The increase in desulfurization plants has slowed down due partly to the successful reduction of ambient SO_2 concentrations, and partly to oversupply of by-products, and to the recent economic depression. The growing usage of coal for utility boilers, however, will necessitate more FGD plants.

Recent efforts for air pollution control have been concentrated on NOx abatement. In addition to combustion modification for numerous flue gas sources, over 60 commercial plants for selective catalytic reduction of NOx in flue gases have been put into operation. Selective noncatalytic reduction (thermal de-NOx) has also been applied commercially or for large scale tests.

The status of technologies, problems, and economics of SO_2 and NOx removal in Japan will be described.

SO_2 AND NOx REMOVAL TECHNOLOGY IN JAPAN

1 FGD PROCESSES AND PLANTS

1.1 Major Processes

Table 1 lists major constructors of FGD plants and numbers and capacities of plants operational at the end of 1978. The plants totaled more than 500 and their combined capacity reached 88,000,000 Nm^3/hr (equivalent to 29,000 MW). About half of the capacity is accounted for by utility boilers (mostly oil-fired) and the rest by industrial boilers, iron-ore sintering machines, nonferrous metal industry, sulfuric acid plants, etc.

About 50% of the plants, in terms of capacity, use the wet lime/limestone process to by-produce gypsum, 16% the indirect lime/limestone process (double alkali type) to by-produce gypsum, 13% the regenerable process to by-produce sulfuric acid, ammonium sulfate and elemental sulfur, and 24% sodium scrubbing to by-produce sodium sulfite or sulfate. The average plant capacity is 443,000 Nm^3/hr for the wet lime/limestone, 279,000 Nm^3/hr for the indirect lime/limestone, 369,000 Nm^3/hr for the regenerable processes, and 59,600 Nm^3/hr for the sodium scrubbing process. About 80% of the sodium scrubbing plants by-produce sodium sulfite for paper mills and the rest oxidize the sulfite by air bubbling to sulfate, which is either used in the glass industry or purged in wastewater.

In addition to the 335 sodium scrubbing plants listed in Table 1, there are about 500 smaller ones operated commercially with an average capacity of about 20,000 Nm^3/hr.

1.2 Status of FGD Plants by Power Companies

Table 2 lists power companies and their capacities of steam power generation and FGD. The nine major power companies (Nos. 1 to 9 in the table) have produced about 70% of total steam power using mainly oil with some LNG and a little coal. Electric Power Development Co. (EPDC, No. 10 in the table) which was established by the nine major companies and the Central Government has been the major consumer of coal for power generation. Other power suppliers have relatively small capacities, burning mainly oil.

Among the major power companies, Tokyo Electric, Kansai Electric and Chubu Electric have the largest power generation capacities (A) and relatively small FGD capacities (B), with a B/A ratio of only 1-8%. Those companies prefer clean fuel such as LNG and low-sulfur oil to FGD, because they have power plants near large

Table 1 NUMBERS AND CAPACITIES (1,000 Nm^3/hr) OF FGD PLANTS BY MAJOR CONSTRUCTORS (OPERATIONAL AT END 1978)

Plant constructor	Wet lime limestone	Indirect lime limestone	H_2SO_4, S $(NH_4)_2SO_4$	Na_2SO_3 Na_2SO_4	Total
Mitsubishi Heavy Industries (MHI)	34 (19,020)			3 (292)	37 (19,312)
Ishikawajima H.I. (IHI)	17 (4,445)			79 (4,351)	96 (8,796)
Hitachi, Ltd.	13 (6,940)*		2 (590)	15 (603)	30 (8,133)
Mitsubishi Kakoki (MKK)	2 (256)		13 (6,478)**	41 (913)	56 (7,643)
Kawasaki Heavy Industries	4 (756)	6 (5,450)		7 (256)	17 (6,380)
Tsukishima Kikai (TSK)	1	4 (398)	1 (88)	40 (4,042)	45 (4,608)
Chiyoda Chemical Eng. & Construc.		15 (4,585)			15 (4,585)
Oji Koei				57 (4,280)	57 (4,280)
Sumitomo Metal - Fujikasui	7 (3,954)			6 (270)	13 (4,224)
Kurabo Engineering		5 (603)	1 (18)	106 (3,751)	112 (4,372)
Mitsui Miike-Chemico	4 (2,744)		1 (500)		5 (3,244)
Ebara Manufacturing		11 (1,914)		10 (1,167)	21 (3,081)
Kobe Steel	6 (2,475)				6 (2,475)
Nippon Kokan (NKK)	3 (245)	1 (150)	2 (1,990)	6 (62)	12 (2,447)
Kureha Chemical				8 (1,431)	8 (1,431)
Showa Denko				5 (1,372)	5 (1,372)
Gadelius				8 (1,291)	8 (1,291)
Sumitomo (SCEC)-Wellman			6 (1,288)		6 (1,288)
Nippon Steel	2 (1,200)				2 (1,200)
Mitsui Metal Engineering	4 (1,006)		2 (130)		6 (1,136)
Dowa Engineering		8 (666)			8 (666)
JGC	1 (330)		1 (125)		2 (455)
Ube Industries			2 (220)		2 (220)
Niigata Engineering		1 (185)			1 (185)
Mitsui Engineering				1 (160)	1 (160)
Total	98 (43,371)	51 (13,951)	31 (11,427)	335 (19,961)	515 (88,710)

* Babcock - Hitachi

** Wellman - MKK

Table 2 CAPACITIES OF STEAM POWER GENERATION AND FGD OF POWER COMPANIES

		Power generation (MW)			FGD (MW)			
No.	Power company	Existing	Under construction*	Total (A)	Existing	Under construction*	Total (B)	B/A (%)
1	Hokkaido	1,270	1,225	2,495	0	525	525	21.0
2	Tohoku	3,925	1,200	5,125	900	0	900	17.6
3	Tokyo	19,167	4,400	23,567	283	0	0	1.2
4	Chubu	9,933	3,800	13,733	970	0	970	7.1
5	Hokuriku	1,412	1,000	2,412	600	500	1,100	45.6
6	Kansai	10,672	1,200	11,872	930	0	930	7.8
7	Chugoku	3,777	1,800	5,777	1,350	700	2,050	36.8
8	Shikoku	2,687	450	3,137	900	0	900	12.5
9	Kyushu	4,500	2,700	7,200	1,626	0	1,626	22.6
10	EPDC	1,430	1,000	2,430	1,280	1,000	2,280	93.8
11	Niigata	350	350	700	175	175	350	50.0
12	Showa	550	0	550	400	0	400	72.7
13	Toyama	750	0	750	250	0	250	33.3
14	Mizushima	462	0	462	156	0	156	33.8
15	Sumitomo	368	250	618	156	0	156	25.2
16	Sakata	0	700	700	700	0	700	100.0
17	Fukui	0	250	250	0	250	250	100.0
18	Others	5,512	375	5,887	0	0	0	0.0
	Total	66,775	20,700	87,475	10,676	3,150	13,826	15.8

* Including those decided to be constructed.

cities such as Tokyo, Osaka and Nagoya and have anticipated the SOx and NOx regulations there to become too stringent to be met by FGD and combustion modification when high-sulfur fuels are used. On the other hand, Hokuriku Electric, Chugoku Electric, EPDC and some of the smaller companies, with power plants relatively distant from large cities, have larger B/A ratios. FGD plants of power companies are listed in Tables 3 and 4. All of the plants by-produce gypsum except the three that by-produce sulfuric acid.

The recent oversupply of gypsum and other FGD by-products and the relatively low cost of low-sulfur fuels due to economic depression has discouraged construction of additional FGD plants. Four plants for coal-fired utility boilers, 175 MW (existing), 250 MW, 500 MW and 500 MW (new), will be completed between 1979 and 1981, all using the limestone-gypsum process.

2 OPERATION OF FGD PLANTS

2.1 Wet Lime/Limestone Process

Table 5 shows operation data of major wet lime/limestone process plants. Those plants by-produce salable gypsum except the Omuta plant, Mitsui Aluminum, which by-produces a throw-away sludge. For the production of gypsum, a calcium sulfite slurry discharged from a scrubber at a pH of 6-7 is treated to lower the pH to about 4 and then is air-oxidized. For the pH adjustment, an additional scrubber is used at the Owase plant, Chubu Electric, and the Takasago plant, EPDC, while sulfuric acid is used at other plants.

SO_2 removal efficiency ranges from 90 to 98%, and power required for a total FGD system from 2.0 to 3.5% of the power generated. The power requirement is larger for a scrubber with a venturi because of the larger pressure drop of the gas in the scrubber to attain high removal efficiencies for SO_2 and dust.

Wastewater is purged at a rate of 3-30 tons/hr or 8-60 kg/MWhr, primarily to prevent the accumulation of chlorine in the scrubber liquor because chlorine increases corrosion.

The Omuta plant, Mitsui Aluminum, has been operated with a low oxidation ratio preventing the formation of gypsum while in other plants a considerable amount of gypsum crystals are added as seeds to a circulating slurry. Scaling can be minimized in either way. None of the plants, except the Tamashima plant, Chugoku Electric, has a stand-by scrubber.

Some of the plants encountered problems at start-up but most of the problems were solved in a few months. All of the plants have since attained an operability better than 97%. Operability means an FGD plant's operating hours per cent of the

Table 3 FGD PLANTS OF POWER COMPANIES (I) (FOR OIL-FIRED BOILERS)

Power company	Power station	Boiler No.	Boiler MW	FGD MW	Process developer	Absorbent, precipitant	By-product	Year of completion
Tohoku	Shinsendai	2	600	150	Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	Gypsum	1974
"	Hachinohe	4	250	125	Mitsubishi H.I.	CaO	"	1974
"	Niigata	4	250	125	Wellman-MKK	Na_2SO_3	H_2SO_4	1976
"	Niigata H.	1	600	150	Mitsubishi H.I.	$CaCO_3$	Gypsum	1976
"	Akita	3	350	350	Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	"	1977
Tokyo	Kashima	3	600	150	Hitachi-Tokyo	Carbon, $CaCO_3$	"	1972
"	Yokosuka	1	265	133	Mitsubishi H.I.	$CaCO_3$	"	1974
Chubu	Nishinagoya	1	220	220	Wellman-MKK	Na_2SO_3	H_2SO_4	1973
"	Owase	1	375	375	Mitsubishi H.I.	CaO	Gypsum	1976
"	"	2	375	375	"	"	"	1976
Hokuriku	Toyama	1	500	250	Chiyoda	H_2SO_4, $CaCO_3$	"	1974
"	Fukui	1	350	350	"	"	"	1975
"	Nanao	1	500	500	Not decided		"	1978
Kansai	Sakai	8	250	63	Sumitomo H.I.	Carbon	H_2SO_4	1972
"	Amagasaki	2	156	35	Mitsubishi H.I.	CaO	Gypsum	1973
				121	"	"	"	1975
"	"	1	156	156	"	"	"	1976
"	Osaka	3	156	156	Babcock-Hitachi	$CaCO_3$	"	1975
"	"	2	156	156	"	"	"	1975
"	"	4	156	156	"	"	"	1976
"	Kainan	4	600	150	Mitsubishi H.I.	CaO	"	1974
Chugoku	Mizushima	2	156	100	Babcock-Hitachi	$CaCO_3$	"	1974
"	Tamashima	3	500	500	"	"	"	1975
"	"	2	350	350	"	"	"	1976
"	Shimonoseki	2	400	400	Mitsubishi H.I.	"	"	1976

Table 4 FGD PLANTS OF POWER COMPANIES (II)

Power company	Power station	Boiler No.	Boiler MW	FGD MW	Process developer	Absorbent, precipitant	By-product	Year of completion
Shikoku	Anan	3	450	450	Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	Gypsum	1975
"	Sakaide	3	450	450	"	"	"	1975
Kyushu	Karita	2	375	188	Mitsubishi H.I.	CaO	"	1974
"	Karatsu	2	375	188	"	$CaCO_3$	"	1976
"	"	3	500	250	"	"	"	1976
"	Ainoura	1	375	250	"	"	"	1976
"	"	2	500	250	"	"	"	1976
"	Buzen	1	500	250	Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	"	1977
"	"	2	500	250	"	"	"	1978
EPDC	Takasago	1	250*	250	Mitsui-Chemico	$CaCO_3$	"	1975
"		2	250*	250	"	"	"	1976
"	Isogo	1	265*	265	Chemico-IHI	"	"	1976
"		2	265*	265	"	"	"	1976
"	Takehara	1	250*	250	Babcock-Hitachi	"	"	1977
Niigata	Niigata	1	350	175	MHI	"	"	1975
Showa	Ichihara	1	150	150	Showa Denko	Na_2SO_3, $CaCO_3$	"	1973
"	"	5	250	250	Babcock-Hitachi	$CaCO_3$	"	1976
Toyama	Toyama	1	250	250	Chiyoda	H_2SO_4, $CaCO_3$	"	1975
Mizushima	Mizushima	5	156	156	Mitsubishi H.I.	CaO	"	1975
Sumitomo	Niihama	3	156	156	IHI	$CaCO_3$	"	1975
Sakata	Sakata	1	350	350	Mitsubishi H.I.	$CaCO_3$	"	1976
"	"	2	350	350	"	"	"	1977
Fukui	Fukui	1	250	250	Not decided	"	"	1979

* Coal-fired boilers. Others are for oil-fired boilers.

Table 5 OPERATION DATA OF MAJOR LIME/LIMESTONE PROCESS PLANTS

Process	Lime scrubbing		Limestone scrubbing					
Process developer	Mitsubishi H.I.	Chemico-Mitsui	Mitsubishi H.I.	Babcock-Hitachi	Babcock-Hitachi	Mitsui-Chemico	IHI-Chemico	Sumitomo-Fujikasui
Plant owner	Chubu Electric	Mitsui Aluminum	Kyushu Electric	Chugoku Electric	Electric Power D.C.	Electric Power D.C.	Electric Power D.C.	Sumitomo Metal
Plant site	Owase	Omuta[a)]	Karatsu	Tamashima	Takehara	Takasago	Isogo	Kashima
Fuel	Oil	Coal	Oil	Oil	Coal	Coal	Coal	Coke
FGD capacity (MW)	375	156	250	500	250	250	265	(630)[b)]
FGD capacity (1,000 Nm^3/hr)	1,200	510	730	1,480	850	850	900	2,000
Inlet SO_2 (ppm)	1,600	2,300	530	1,460	1,550	1,500	450	400-600
Inlet dust (mg/Nm^3)	15	630	25	40	600	100	1,500	100-200
Inlet gas temperature (°C)	150	135	150	140	150	140	170	150
CaO/SO_2 stoichiometry	1.0	1.05			1.05	1.0	1.0-1.05	1.05
Number of scrubbers in parallel	2	2	1	3+1[c)]	1	1	1	2
Prescrubber (first scrubber)								
Type	Spray	Venturi	Spray	Venturi	Venturi	Venturi	Venturi	PP[d)]
L/G ($liters/Nm^3$)	2	5-8	2	10	2.4	6	7	7
Scrubber (second scrubber)								
Type	Packed	Venturi	Packed	PP[d)]	PP[d)]	Venturi	Venturi	PP[d)]
Slurry pH	6.5-7	7.5	6.2	6.6	6-6.5	6.2	5-6	6.7
Slurry concentration (%)	10	5		12.5	9	5-6	7	7-8
L/G ($liters/Nm^3$)	7	5-8	12	10	7	6	7	8
Gas velocity (m/sec)	3.4		3.0				3.0	4.4
Outlet SO_2 (ppm)	120	220	50	60	100	100	10	30
Outlet dust (mg/Nm^3)	8	40	6		30	50	50	20
SO_2 removal efficiency (%)	93	90	90	96	98	93	93	93-95
Mist eliminator type	CE[e)]	Chevron	Chevron	Pwf[f)]	Pwf[f)]	Chevron	Chevron	Impinger
Pressure drop (mmH_2O) Prescrubber	30	200	60	225	230	150	150	180
Scrubber	150		35	505	375	150	150	130
Mist eliminator	25	30	65	25	10	25	50	30
Total system	375		185	1,080	950	525	600	530
Wastewater purged (t/hr)	3			4	15	10	15	30
Energy requirement (design)								
Pump (kW)		1,320		4,100	1,400	2,800	2,000	4,400
Fan (kW)		2,000		11,500	5,500	4,200	4,800	8,170
Total FGD system (kW)	7,500	3,867	5,070	17,600	8,200	8,000	7,800	13,230
Per cent of power generated	2.0	2.5	2.0	3.5	3.3	3.2	2.9	2.1
Operability (%)	98, 99	100	99.7	97.4	97.4	97, 98	100	100

a) Throw-away process. Others by-produce gypsum. b) Iron-ore sintering plant. Others are utility boilers.
c) Stand-by. d) Perforated plate. e) Chevron and Euroform. f) Pipe with fin.

desired operating hours of the gas source in a year, as will be desctibed in detail in Section 2.4.

Figure 1 shows the relationship of the operability and inlet SO_2 concentration of major wet lime/limestone process plants. The operability is generally lower with higher inlet SO_2 concentrations, indicating a greater scaling tendency. The difference in operability between the plants for oil-fired and coal-fired boilers is little and may not be significant. Two plants for industrial boilers have 100% operability in spite of high inlet SO_2 concentrations, possibly because of easy operation control due to the stable boiler load.

2.2 Indirect and Modified Lime/Limestone Process

Table 6 shows the operation data of major plants by-producing gypsum by indirect lime/limestone processes (double alkali type) and the Kobe Steel and Kawasaki processes (modified lime/limestone process) which use-- in addition to lime-- calcium chloride and magnesia, respectively. The pH of the absorbing liquor of the processes ranges from 1.0 (Chiyoda) to 6.8 (Showa Denko). For processes which use a liquor with a pH below 4.0 (Chiyoda, Dowa, and Kurabo), oxidation is carried out with the liquor and proceeds more rapidly than with the calcium sulfite in other processes. The SO_2 removal efficiency ranges from 90 to 96%.

Figure 2 shows the relationship of pH and plant performance. The lower the pH, the larger the L/C ratio and power consumption, and the higher the operability. The higher operability may be due to the lesser tendency of scaling.

2.3 Regenerable Processes

Table 7 shows operation data of major regenerable process plants. A large ammonia scrubbing plant by the Nippon Kokan process has been operated with 100% operability (Section 3.2). Wellman-Lord process plants have been operated smoothly but require extensive wastewater treatment including ozone oxidation to decompose polythionates such as $Na_2S_2O_5$, which form mainly at the heating step of the absorbing liquor. Polythionates form also in other wet processes even though in small amounts and might necessitate treatment when wastewater regulations are tightened.

The Chemico-Mitsui magnesium scrubbing plant of Idemitsu Kosan encountered problems for over 1 year after its start-up in 1975. The problems have since been solved through improvements of the process by Idemitsu Kosan and Mitsui Miike. Normally no wastewater is purged from the system.

Sodium scrubbing to by-produce sodium sulfite for paper mills (such as the Kureha process in the Table) is most simple and requires the least energy, but

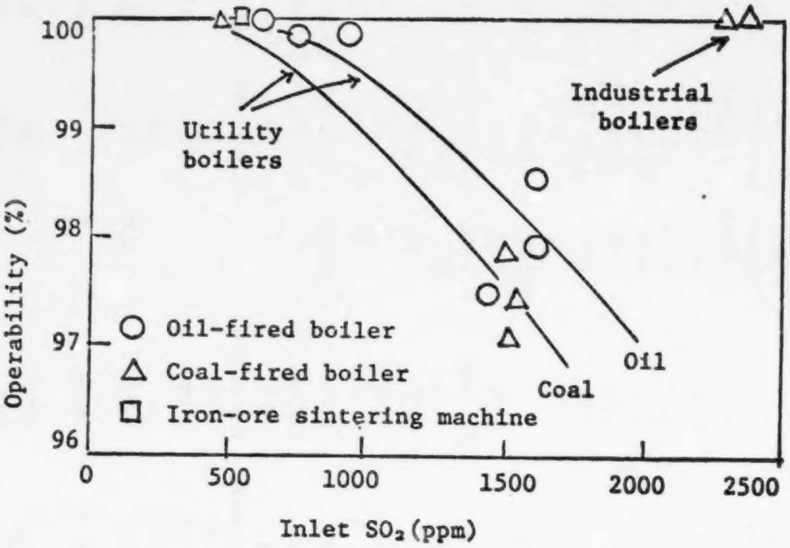

Figure 1 Relationship of inlet SO_2 concentration and operability of FGD (lime/limestone process)

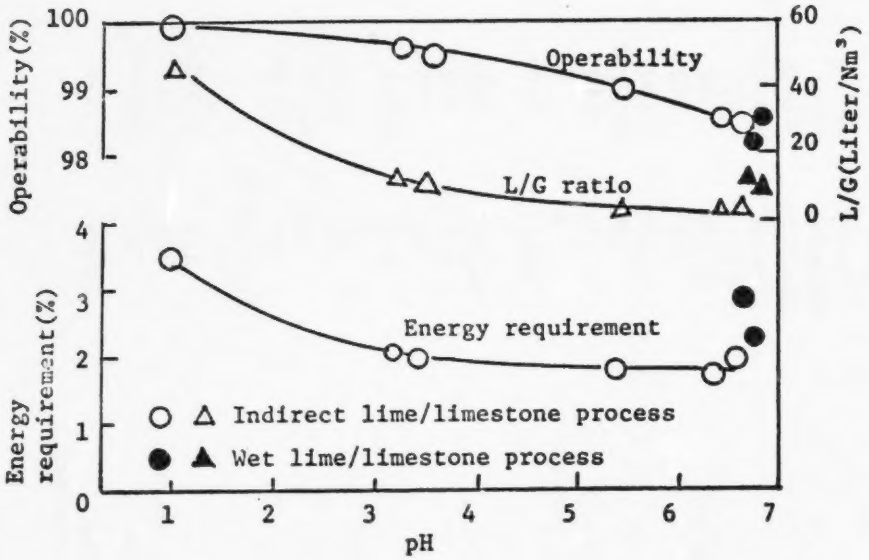

Figure 2 pH of scrubber liquor and performance of FGD

Table 6 OPERATION DATA OF INDIRECT AND MODIFIED LIME/LIMESTONE PROCESS PLANTS

Process developer	Kureha-Kawasaki	Showa Denko	Chiyoda	Dowa	Kurabo	Nippon Kokan	Kobe Steel	Kawasaki H.I.
Absorbent	NaOH	NaOH	H_2SO_4	$Al_2(SO_4)_3$	NH_3	NH_3	CaO-$CaCl_2$	$Mg(OH)_2$
Precipitant	$CaCO_3$	$CaCO_3$	$CaCO_3$	$CaCO_3$	CaO	CaO		CaO, $CaCO_3$
Plant owner	Shikoku Electric	Showa Denko	Hokuriku Electric	Naikai Salt		Nippon Kokan	Nakayama Steel	Unitika
Plant site	Sakaide	Ichihara	Fukui	Tamano		Keihin	Funamachi	Okazaki
Fuel	Oil	Oil	Oil	Oil	Oil	Coke[a]	Coke[a]	Oil
FGD capacity (1,000 Nm^3/hr)	1,270	500	980	72	115	150	375	200
FGD capacity (MW)	450	150	350	(25)	(40)	(50)	(125)	68
Inlet SO_2 (ppm)	1,270	1,400	1,800	1,500	1,480	350	150-250	1,400
Inlet dust (mg/Nm^3)	20	100-200	30	200	150	80	300-400	200
Inlet gas temperature (°C)	135	140	140	170	170	120	140-155	170
Number of scrubbers in parallel	2	4	1	1	1	1	1	1
Prescrubber type	None	None	Venturi	Spray		Spray	Spray	None
L/G (liters/Nm^3)				2.5	1.4	1.0	3.5	
Scrubber type	Packed	VC[b]	Packed	Packed	Packed	Screen	Venturi	MV[c]
Liquor pH	6.2	6.8	1	3.5	3.8	6.0	5-5.5	5-6
Concentration	20	25	1-2		10	30	30+6[d]	
L/G (liters/Nm^3)	1.2	0.5-1	40-60	10	8	2	3	6
Gas velocity (m/sec.)	2			1.2	2	2	3	3
Outlet SO_2 (ppm)	70	50-90	80	100	130	10-20	15-25	Below 140
Outlet dust (mg/Nm^3)	10	Below 50		50	Below 50		40-50	Below 100
SO_2 removal efficiency (%)	95	93-96	96	93	91		90	Above 90
Mist eliminator type	Terellette		Euroform	Wire mesh	Euroform		Euroform	Louver
Pressure drop (mmH_2O) Prescrubber				10-20	130		15	
Scrubber	115	300-500		70-80	120		120	115
Mist eliminator	25	30-50		20-30			35	50
Total system	310	400-700	680	170	300	300	220	230
Wastewater purged (t/hr)	None	4-6	7-24	0.3	None	2	None	None
Energy requirements (Design)								
Pump (kW)			5,300	150	470		1,200	430
Fan (kW)			5,500	300	190		1,000	370
Total system (kW)	7,900	3,500	12,300	580	880		2,600	1,160
Per cent of power generated	1.8	2.3	3.5	2.3	2.2		2.1	1.7
Operability (%)	98.6	98.7	100	99.6	99.3	Above 95	98.1	99

a) Iron-ore sintering plant b) Vertical cone c) Multi-venturi d) 30% $CaCl_2$ + 6% CaO

Table 7 OPERATION DATA OF REGENERABLE PROCESS PLANTS

Process developer	Nippon Kokan	Wellman-MKK	Wellman-SCEC	Mitsui-Chemico	Kureha	Shell[c)]	TEPCO-Hitachi
Absorbent	NH_3	NaOH	NaOH	MgO	NaOH	CuO	Carbon
By-product	$(NH_4)_2SO_4$	$SO_2 \rightarrow H_2SO_4$	$SO_2 \rightarrow H_2SO_4$	$SO_2 \rightarrow S$	Na_2SO_3	$SO_2 \rightarrow S$	$H_2SO_4 \rightarrow CaSO_4$
Plant owner	Nippon Kokan	Chubu Electric	Sumitomo Chemical	Idemitsu Kosan	Mitsui Toatsu	Showa Y.S.	Tokyo Electric
Plant site	Ogishima	Nishinagoya	Sodegaura	Chiba	Nagoya	Yokkaichi	Kashima
Fuel	Coke[a)]	Oil	Oil	Oil	Oil	Oil	Oil
FGD capacity (1,000 Nm^3/hr)	1,120	620	370	460[b)]	190	116	420
FGD capacity (MW)	(380)	220	130	(160)	65	38	150
Inlet SO_2 (ppm)	350	1,600	1,500	2,850	1,400	1,250	150
Inlet dust (mg/Nm^3)	50		100		200-300	Below 50	
Inlet gas temperature (°C)	120	140	160	185	170		130
Number of scrubbers in parallel	2	1		1		1[d)]	
Prescrubber type	Spray			Venturi			
L/G (liters/Nm^3)	1.0						
Scrubber type	Screen	Sieve tray		Venturi	Packed	pp[e)]	Packed
Liquor pH	6.0				6.5		
L/G (liters/Nm^3)	1.0	0.6			1.2		
Gas velocity (m/sec)	1.6	1.8			Below 2.0		
Outlet SO_2 (ppm)	10-20	120	Below 150	120	6	125	30
Outlet dust (mg/Nm^3)	10	35	Below 50			Below 50	
SO_2 removal efficiency (%)	94-97	92	Over 90	95		90	80
Mist eliminator type	Wet EP			Chemico			
Pressure drop (mmH_2O) Scrubber		400			165	200	630
Pressure drop (mmH_2O) Mist eliminator		50			40		
Pressure drop (mmH_2O) Total system	250	550		500[f)]	250	400-500[f)]	870
Wastewater purged (t/hr)	10	4	Some	0.1	Some	Minor	13
Energy requirements (Design)							
Pump (kW)		840		1,960		140	280
Fan (kW)		2,350		3,400		730	2,700
Total FGD system (kW)					560	870[f)]	3,245
Per cent of power generated		1.5[h)]		3.4[h)]	0.9	2.3[f,g)]	2.2
Operability (%)	100	97.8	Over 95	98	100	88[i)]	92[i)]

a) Iron ore sintering plant b) From oil burner and Claus furnace c) Dry process removing up to 70% of NOx simultaneously d) Two reactors are used alternately for SO_2 absorption and regeneration e) Parallel passage reactor f) Excluding Claus furnace g) Including energy for steam h) For pump and fan i) FGD plant was shut down for inspection demand for the sulfite is limited.

The Shell process plant of SYS which had a low operability for a few years achieved ten months continuous operation recently. Although the process seems costly, the capability of simultaneous removal of NOx may compensate for the disadvantage.

The Kashima plant, Tokyo Electric, using carbon absorption and water wash, has been operated for 6 years without appreciable problems. Carbon consumption was proved low (about 2% yearly) owing to the use of a fixed bed and to regeneration by water wash.

2.4 Operability

As shown in Table 8, most FGD plants have an operability of over 97%. Operability means FGD operation hours per cent of the desired operation hours of FGD, vis. the scheduled gas source operation hours less the hours of shutdown caused by trouble with the gas source. For oil-fired boilers, when an FGD plant has to be shut down due to its own trouble, boiler operation is continued by switching immediately to a low-sulfur oil. For coal-fired boilers and sintering machines, it can happen that the gas source has to be shut down due to FGD trouble. In such a case the true operability is less than the FGD operation hours per cent of the gas source operation hours, as shown for EPDC's Takasago plant in Table 8.

Three plants in the table have an operability of 100% but this does not mean the plants have no problems at all. Minor problems can be solved without interrupting operation. For example, most of the plants have a stand-by pump to replace a pump during FGD operation. With corrosion, scaling and plugging under a certain level, the FGD plant can be operated continuously until the scheduled shutdown of the gas source, when it can be repaired. Normally boilers are operated continuously for 11 months and then undergo one month's shutdown for annual maintenance, while iron-ore sintering machines are operated continuously for 2 or 3 months and shut down for several days for maintenance.

2.5 Labor Requirement

Most larger FGD plants are operated by 2-4 persons per shift who also carry out minor maintenance work. For annual maintenance to solve serious problems of an FGD system, a skilled maintenance staff who takes care of the whole power plant (or steel plant, etc.) looks after the FGD system.

Typical labor requirements are shown in Table 9. The Fukui plant, Hokuriku Electric, by the Chiyoda process shows the least labor needs indicating trouble-free operation. On the other hand, the Sakaide plant, Shikoku Electric, shows the largest man-hour requirements, although the operability is relatively high, presumably

Table 8 OPERATION HOURS AND FGD OPERABILITY IN RECENT ONE YEAR

Plant owner	Plant site	Gas source	Process	FGD capacity (MW)	Operation (hr/year) Boiler (A)	Operation (hr/year) FGD (B)	Operability (B/A)x 100	Inlet SO_2 (ppm)	Year completed
Lime-Limestone process									
Chubu Electric	Owase (1)	UB[a]	MHI	375	7,320	7,171	98.0	1,600	1976
"	" (2)	"	"	375	7,565	7,485	98.9	1,600	1976
Kyushu Electric	Kanda	"	"	188	7,420	7,390	99.6	800	1974
"	Karatsu	"	"	240	7,271	7,246	99.7	530	1976
Electric P.D.C.	Takasago[b]	"	Mitsui-Chemico	250	8,180	8,010	97.9[e]	1,500	1975
Mitsui Aluminum	Omuta	IB[c]	Chemico-Mitsui	156	8,244	8,232	99.9	2,300	1972
"	"	"	Mitsui-Chemico	175	8,040	8,040	100.0	2,300	1975
Sumitomo Metal	Kashima	SM[d]	Sumitomo-Fujikasui	(330)	8,285	8,285	100.0	500	1976
Indirect or modified lime-limestone process									
Shikoku Electric	Sakaide	UB	Kureha-Kawasaki	450	7,441	7,336	98.6	1,270	1975
Hokuriku Electric	Fukui	"	Chiyoda	350	7,044	7,044	100.0	1,800	1975
Showa Denko	Ichihara	IB	Showa Denko	150	7,885	7,775	98.6	1,400	1973
Unitika	Okazaki	"	Kawasaki	67	8,232	8,160	99.1	1,400	1975
Naikai Salt	Tamano	"	Dowa	28	8,001	7,969	99.6	1,500	1976
Nippon Kokan	Keihin	SM	Nippon Kokan	(50)	8,202	8,098	98.7	350	1972
Nakayama Steel	Kobe	"	Kobe Steel	(125)	8,419	8,259	98.1	200	1976
Regenerable process									
Idemitsu Kosan	Chiba	IB	Chemico-Mitsui	170	8,016	7,887	98.4	2,850	1975
Chubu Electric	Nishinagoya	UB	Wellman-MKK	220	7,247	7,090	97.8	1,500	1973
Showa Y.S.	Yokkaichi	IB	Shell	40	7,656	6,720	87.8[f]	1,250	1973
Nippon Kokan	Fukuyama	SM	Nippon Kokan	(253)	4,263[g]	4,263[g]	100.0	350	1976

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a) Utility boiler.
b) Coal-fired; other boilers are oil-fired.
c) Industrial boiler.
d) Sintering machine.
e) True operability is 97.0% because boiler stopped for hours due to FGD trouble.
f) FDG plant was shut down for inspection.
g) Due to production control.

Table 9 LABOR REQUIREMENTS OF FGD PLANTS (recent one year)

Plant owner (plant site)	Process (Absorbent)	FGD capacity (MW)	Operation personnel (man-hours/year)				Operability (%)
			Skilled	Unskilled	Maintenance	Total	
Shikoku Electric (Sakaide)	Kureha-Kawasaki (Na_2SO_3-$CaCO_3$)	450	33,000	16,000	19,900	68,900	98.6
Mitsui Aluminum[a] (Omuta)	Chemico-Mitsui (Carbide lime)	156	8,040	8,040	15,360	31,440	100
Chugoku Electric[b] (Tamashima)	Babcock-Hitachi ($CaCO_3$)	500	17,520	17,520	not clear		97.4
Chubu Electric (Nishinagoya)	Wellman-MKK (Na_2SO_3)	220	17,000		14,000	31,000	97.8
Chubu Electric (Owase-Mita)	CaO	375	17,000		14,000	31,000	98.5
Hokuriku Electric (Fukui)	Chiyoda (H_2SO_4-$CaCO_3$)	350	20,800		3,300	24,100	100

a) Coal-fired; others are oil-fired.

b) Has three scrubbers and a stand-by scrubber; others have no stand-by.

because the process is not simple.

3 NEW FGD TECHNOLOGY

3.1 Chiyoda Jet-Bubbling Limestone-Gypsum Process

Chiyoda Chemical Engineering & Construction Co. has developed a new FGD process using a multi-function jet bubbling reactor which serves as absorber, a limestone reactor, an oxidizer and a gypsum crystallizer requiring no slurry circulation pump.2) Following pilot plant tests in Japan, a prototype unit with a capacity of treating 85,000 nm^3/hr of flue gas from a coal-fired boiler (23 MW equivalent) was constructed at Gulf Power Company's Scholz Station in Florida, U. S.3) The plant went into operation in August 1978 and has been operated since at nearly 100% operability without any scaling problem removing 90-93% of SO_2 with 250 mmH_2O pressure drop in the reactor and 100-101% stoichiometry of limestone. Details of the operation will be given by EPRI at the present symposium.

3.2 Sodium Limestone Process With Electrolytic Cell

The Buzen plant, Kyushu Electric, with a capacity of treating half the flue gas from a 500 MW oil-fired boiler by the Kureha-Kawasaki sodium limestone process, has used an electrolytic cell originally developed by Ionics, U. S., to decompose by-product Na_2SO_4 to NaOH and N_2SO_4. About 800 kg/hr of Na_2SO_4 formed in the FGD system is sent to the cell in a 20% solution after being treated by NaOH and Na_2CO_3 to remove impurities. The by-produced NaOH is sent to the scrubber system and the H_2SO_4 is sent to an oxidizer of calcium sulfite to promote the oxidation. No wastewater is purged from the system.

The plant has been operated since November 1977 and the cell had a corrosion problem at the beginning. The problem has been reduced by changing the construction materials of the cell.

3.3 Nippon Kokan Ammonia Scrubbing Process 1)

Nippon Kokan (NKK) recently completed two large FGD plants to treat flue gas from iron-ore sintering machines to produce ammonium sulfate utilizing ammonia in coke oven gas. A flow sheet of one of them, the Ogishima plant, is shown in Figure 3. Flue gas (1,120,000 Nm^3/hr) is treated with ammonium sulfite liquor to remove over 95% of SO_2. The liquor discharged from the scrubbers (two in parallel) containing ammonium bisulfite is contacted with coke oven gas to absorb ammonia. A portion of the resulting ammonium sulfite liquor is oxidized to produce ammonium sulfate. The operation parameters are shown in Table 7.

Hydrogen sulfide in the coke oven gas is removed prior to the ammonia absorption.

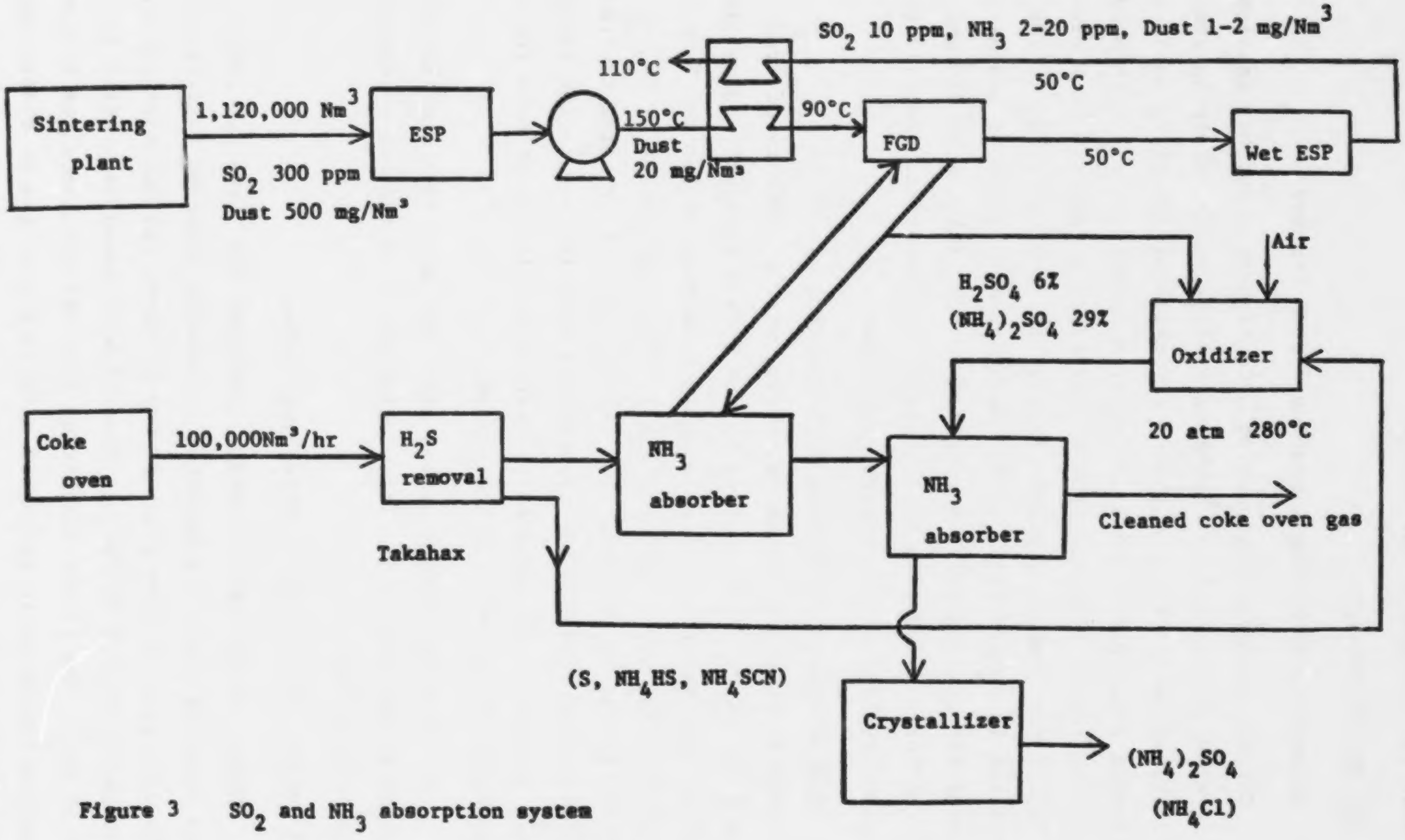
Sintering plant
1,120,000 Nm^3
SO$_2$ 300 ppm
Dust 500 mg/Nm^3
ESP
150°C
Dust
20 mg/Nm^3
110°C
90°C
FGD
50°C
Wet ESP
50°C
SO$_2$ 10 ppm, NH$_3$ 2-20 ppm, Dust 1-2 mg/Nm^3
Air
H$_2$SO$_4$ 6%
(NH$_4$)$_2$SO$_4$ 29%
Oxidizer
20 atm 280°C
Coke oven
100,000Nm^3/hr
H$_2$S removal
Takahax
NH$_3$ absorber
NH$_3$ absorber
Cleaned coke oven gas
(S, NH$_4$HS, NH$_4$SCN)
Crystallizer
(NH$_4$)$_2$SO$_4$
(NH$_4$Cl)

Figure 3 SO$_2$ and NH$_3$ absorption system
(Ogishima plant, Nippon Kokan)

By-products from the sulfide removal unit are also oxidized and coverted to ammonium sulfate.

The flue gas leaving the scrubber is passed through wet electrostatic precipitators (eight in parallel) and heated to 110°C by two Ljungstrom type heat exchangers installed in parallel. No plume at all is observed from the stack.

Nippon Kokan has a similar plant at Fukuyama, with a capacity of treating 760,000 Nm^3/hr of flue gas. At this plant, the treated flue gas is heated by conventional oil firing using neither wet electrostatic precipitator nor heat exchanger. An appreciable plume is observed at the stack.

3.4 Ammonia Scrubbing by Ube Industries Process

Ube Industries has developed an ammonia scrubbing process and constructed two commercial units in 1977, each with a capacity of treating 110,000 Nm^3/hr of flue gas from a boiler burning 2% sulfur oil. Ammonium sulfite is by-produced which is highly purified and used for caprolactam production. The total investment cost was nearly 1 billion yen (13,700 yen/kW). An appreciable plume is observed. Tests are in progress to reduce the plume.

3.5 Gas-Gas Heating by Ljungstrom Heat Exchanger

In principle, a gas-gas heat exchanger as shown in Figure 3 is very useful for FGD because it not only saves energy but also reduces the consumption of cooling water. This type of heat exchanger has not been used commercially for FGD because of corrosion and solid deposition within it.

Pilot plant tests were carried out by Gadelius Co. (Japan) jointly with EPDC for a wet limestone FGD system for flue gas from a coal-fired boiler and also with Tohoku Electric for a sodium limestone FGD system for flue gas from an oil-fired boiler. Corrosion-resistant materials and soot blow were used to solve the problems.

Tests for 6,000 hours with oil-fired gas (10,000 Nm^3/hr) containing about 10 mg/Nm^3 of dust showed that a slight deposit less than 0.5 mm in thickness formed in a zone where the soot blow was not effective. Pressure drop did not increase appreciably and the deposit could be removed by water wash.4) Tests with flue gas from coal containing 100-200 mg/Nm^3 of fly ash showed that the solid deposits were much softer than those with oil-fired gas and could be removed by soot blowing. Virtually no corrosion was observed in either case with the elements coated with enamel.

It has been decided to use Ljungstrom type heat exchangers for 3 wet limestone process FGD plants to be completed between 1979 and 1981 for coal-fired boilers.

4 NOx ABATEMENT (5)

4.1 Outline

NOx emissions from stationary sources have been controlled by the emission standards enforced by the Central Government and also by agreements of industry with local governments. The emission standards for boilers are shown in Table 10.

Table 10 NOx EMISSION STANDARDS FOR BOILERS (ppm)

	Capacity (1,000 Nm^3/hr)							
	10-40		40-100		100-500		Over 500	
Fuel	N^a	E^b	N	E	N	E	N	E
Gas	130	150	130	130	100	130	60	130
Oil	150	230	150	190	150	190	130	180
Coal	400	600	400	600	400	480	400	480

a For new boilers

b For existing boilers

Combustion modification has made much progress and has satisfied the emission standards. NOx concentrations in flue gases from some of the utility boilers have been reduced to a very low level--- 40 ppm for gas, 100 ppm for oil, and 200 ppm for coal. Stringent regulations have been applied also to mobile sources. New passenger cars have already met the most stringent NOx emission control, 0.25 g/km. The ambient NO_2 standard, 0.04-0.06 ppm in daily average, however, still cannot be met in large cities and industrial regions. The total mass regulations are to be applied for NOx sources in polluted regions whose daily average NO_2 concentrations exceed 0.06 ppm. In addition, local governments apply very stringent control on new large stationary sources of NOx. Flue gas treatment for NOx removal (flue gas denitrification) is thus needed. Major denitrification processes are shown in Table 11.

Table 11 MAJOR PROCESSES FOR FLUE GAS DENITRIFICATION

Dry processes	Selective catalytic reduction (SCR)	
	Selective noncatalytic reduction (SNR)	
	Simultaneous NOx, SOx removal	Copper oxide process Carbon process Electron beam process
Wet processes	Oxidation absorption	
	Simultaneous NOx, SOx removal	Oxidation reduction Reduction

4.2 Selective Catalytic Reduction (SCR)

Among the denitrification processes, SCR which uses NH_3 and catalyst to reduce NOx to N_2 at 200-450°C has been most popular and adopted at 60 commercial plants because it is simple, can give a high NOx removal efficiency of over 90% and does not give by-products difficult of disposal except the spent catalyst. Major SCR plants of power companies are listed in Table 12.

A small amount of O_2 is needed for the reaction of NH_3 with NOx, which is expressed in the following equation:

$$4NO + 4NH_3 + O_2 = 4N_2 + 6H_2O$$

As the catalyst, base metals such as Fe, V, Cr, Cu, Co, and Mo have been used. As the catalyst carrier (support), Al_2O_3 was used at the beginning because of the large surface area as shown below:6)

$$\gamma\text{-}Al_2O_3 > TiO_2 > ZrO_2 > MgO > \alpha\text{-}Al_2O_3 > SiO_2$$

Since Al_2O_3 reacts with SOx in the gas, and this results in decreases in the surface area and activity, most of the recent catalysts use TiO_2 or its compounds. The SOx resistance of the carrier is as shown below:

$$TiO_2 = SiO_2 > \alpha\text{-}Al_2O_3 > \eta\text{-}Al_2O_3 > \gamma\text{-}Al_2O_3$$

Typical operation data of SCR are shown in Figure 4. For treating flue gas from boilers, usually the gas leaving the boiler economizer at 350-400°C is subjected to SCR. About 90% NOx removal is obtained with about 5-10 ppm of leak NH_3 when an NH_3/NOx mole ratio of 1.0 is used with an SV (space velocity) of 5,000-10,000 hr^{-1}. The gas leaving the reactor is passed through an air preheater for heat recovery.

Table 12 MAJOR NOx REMOVAL PLANTS OF POWER COMPANIES

Plant constructor	Power company	Plant site	Fuel of boiler	NOx removal capacity(MW)	Reactor (Catalyst)	Year of completion
Selective catalytic reduction (SCR, 80-90% removal)						
Hitachi Ltd.	Kansai Electric	Kainan	LSO[a]	100	FB[b]	1977
Hitachi Ltd.	Chubu Electric	Chita	LNG	700 x 2	FB	1977
Hitachi Ltd.	Hokkaido Electric	Tomakomai	Coal	90	MB[c]	1981
MHI	Kyushu Electric	Kokura	LNG	600 x 2	FB	1978
MHI	Chubu Electric	Chita	LSO	700	HC[d]	1980
IHI	Chugoku Electric	Kudamatsu	LSO	700	HC	1979
IHI	Tohoku Electric	Niigata	LSO	600	HC	1981
IHI	Company A		LSO	350	HC	1978
ND[e]	EPDC	Takaehara	Coal	250	ND[e]	1980
Selective noncatalytic reduction (SNR, 45% removal)						
MHI	Chubu Electric	Chita	LSO	375		1977
Combination of SNR and SCR (50 - 70% removal)						
Hitachi Ltd.	Company B		LSO	156	PC[f]	1978
MHI	Company B		LSO	156	PC	1978
IHI	Company B		LSO	156	HC	1978
IHI	Company C		LSO	350	HC	1978

[a] Low-sulfur oil [b] Fixed bed [c] Moving bed (with hot electrostatic precipitator)

[d] Honeycomb catalyst [e] Not decided [f] Parallel plate catalyst

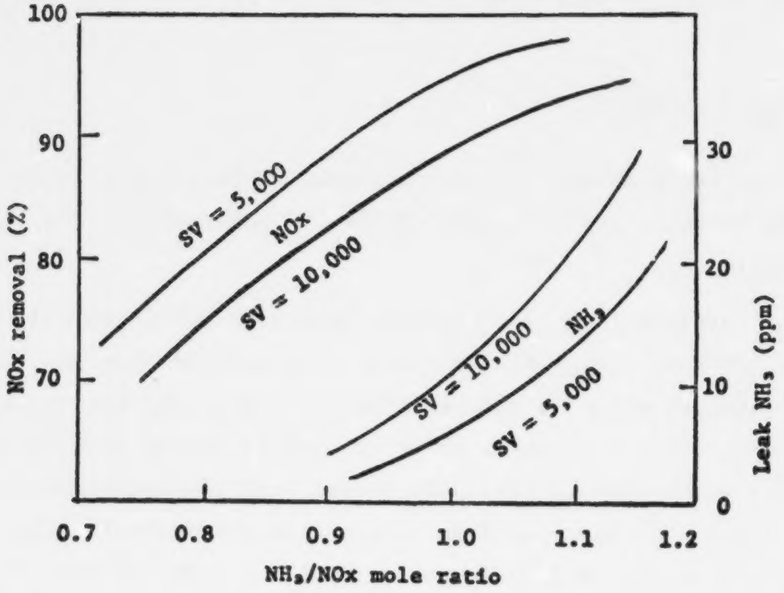

Figure 4 Typical operation data of SCR
(Inlet NOx 150 - 200 ppm)

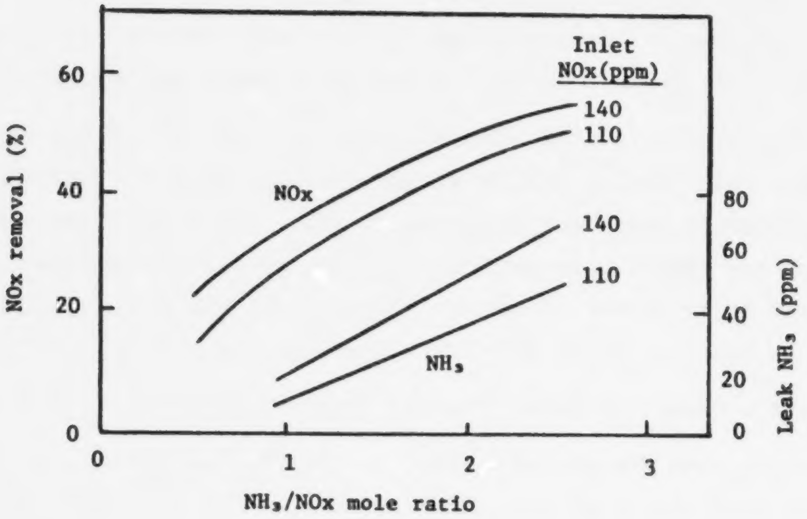

Figure 5 Operation data of SNR at Chita plant,
Chubu Electric (375 MW oil-fired boiler)

One problem with SCR as well as with SNR is the deposit of ammonium bisulfate below about 220°C in the air preheater, which increases the pressure drop and also causes corrosion.

$$NH_3 + SO_3 + H_2O = NH_4HSO_4$$

In order to minimize the formation of the bisulfate, it is preferable to reduce the leak NH_3 below 5 ppm by using less NH_3, viz. 0.85-0.90 mole/mole of NOx, to obtain 80-85% removal efficiency.

Moving bed type and parallel flow type reactors have been developed and used commercially in order to prevent clogging of catalyst with dust in flue gas. The former uses a granular catalyst which is charged from the top of the reactor and is moved down intermittently or continuously while the gas is passed through the catalyst layer in a cross flow. The catalyst discharged from the bottom of the reactor is screened to remove the dust and then returned to the reactor. The reactor usually can treat flue gas containing up to about 0.2 g/Nm^3 of dust.

On the other hand, the parallel flow type reactor uses a fixed bed of a different type of catalyst --- honeycomb, plate, tube, or a parallel passage device. The gas passes through a clearance between the parallel layers of the catalyst and the reactor is expected to be able to handle even a flue gas from a coal-fired boiler containing about 20 g/Nm^3 of dust although the dust may erode the catalyst at a large gas velocity or may deposit on the catalyst at a small gas velocity.

SCR catalysts tend to oxidize a small portion (usually below 5%) of SO_2 in the gas to SO_3. Studies have been made to produce catalysts which cause no oxidation. Catalysts reactive at 150-200°C have been developed to treat low temperature gases. A common problem for the low-temperature catalysts is the deposit of ammonium bisulfate on the catalyst to lower the activity. The activity can be recovered by occasional heating of the catalyst above 350°C to remove the bisulfate.

4.3 Selective Noncatalytic Reduction (SNR) (Thermal De-NOx)

Ammonia rapidly reacts with NOx around 1,000°C to form N_2 and H_2O without catalyst. The important keys to a high removal efficiency by SNR are a good rapid mixing of ammonia with flue gas and a sufficient reaction time (about 0.2 sec.) at a suitable temperature range of 900-1050°C. Those are attained in a laboratory but not easily at an actual plant. Figure 5 shows typical operation data of full scale tests by Chubu Electric with flue gas from a 375 MW oil-fired boiler (0.3% sulfur oil).[7] A total of 15 ammonia injecting nozzles with many holes are placed in two locations in the boiler to cope with the gas temperature fluctuation due to the change of the boiler load.7) Use of 2 moles of NH_3 for each mole of NOx gives 50% removal and also 50 ppm of leak ammonia which causes problems. Routine operation has been carried out using about 1.5 moles of NH_3 to remove about 45% of NOx giving leak NH_3 of about 30 ppm.5)

Mitsui Petrochemical has been using H_2 and NH_3 to remove about 40% of NOx in flue gas from a 40 MW oil-fired boiler by reaction at 700-800°C. A few companies have been testing a combination of SNR and SCR --- ammonia is injected in a boiler at 800-1,000°C and a small amount of parallel-flow type catalyst is placed in a duct at 350-400°C --- to increase NOx removal to 50-70% and to reduce leak NH_3 to about 10 ppm (Table 12). With a larger amount of catalyst, higher removal efficiency is attained but usually the pressure drop becomes too high because of the large gas velocity in the duct.

5 SIMULTANEOUS NOx AND SOx REMOVAL

5.1 Shell Copper Oxide Process

The Yokkaichi plant, SYS, based on the Shell copper oxide process and designed to remove about 90% of SO_2 in flue gas from an oil-fired boiler (40 MW equivalent, 3% sulfur oil), has also removed up to 70% of NOx by adding ammonia to the reactor utilizing the catalytic effect of CuO and $CuSO_4$.8) Pilot plant tests (0.5 MW equivalent) are to be made at Tampa Electric Company's Big Bend Station to remove 90% of both SO_2 and NOx from flue gas from a coal-fired boiler.9)

5.2 Activated Carbon Process 5)

Activated carbon adsorbs SOx and also works as an SCR catalyst, particularly when impregnated with a small amount of metal compound. Flue gas injected with NH_3 is passed through the carbon bed around 220°C with an SV of about 1,000 hr^{-1} for 90% removal of both SO_2 and NOx. A higher temperature increases the NOx removal but decreases the SO_2 removal. SOx is adsorbed by the carbon to form H_2SO_4 and NH_4HSO_4, which are removed by heating the carbon at 350°C in an inert gas produced by incomplete combustion of fuel. Concentrated SO_2 is recovered. Tests have been carried out with pilot plants (0.7 and 2 MW equivalent).

5.3 Electron Beam Process 10)

Flue gas at about 100°C is mixed with NH_3 and exposed to electron beam radiation. About 80% of both SOx and NOx are removed with 2 Mrad of the beam forming fine crystals of ammonium nitrate sulfate double salt which are caught by an electrostatic precipitator for fertilizer use. Tests are in progress at a pilot plant (1 MW equivalent) of Nippon Steel.

5.4 Wet Simultaneous Removal Processes 5)

NO is not readily absorbed on absorbent liquors while NO_2 and N_2O_3 obtained by oxidation of NO are more readily absorbed but the resulting liquor containing nitrate and nitrite is not easily treated. It is possible, however, to reduce absorbed NOx to NH_3 or N_2 by utilizing the reducing effect of SO_2 present in the flue gas.

By the oxidation reduction process for simultaneous removal, NO is oxidized to NO_2, which is absorbed in a limestone slurry containing a catalyst, while by the reduction process, NO is absorbed in a solution of alkaline compound containing EDTA (ethylenediamine tetraacetic acid) and ferrous ion. Although the reactions are complex, forming imidodisulfonate $NH(SO_3M)_2$ and sulfamate NH_2SO_3M (M = Na, K, NH_4, or 1/2Ca) and other imtermediate compounds, the overall reaction may be expressed by the following equations:

$$2NO_2 + 7SO_2 + 7CaCO_3 + 3H_2O = 7CaSO_4 + 2NH_3 + 7CO_2 \qquad 19)$$

$$2NO_2 + 6SO_2 + 6CaCO_3 + O_2 = 6CaSO_4 + N_2 + 6CO_2 \qquad 20)$$

$$2NO + 5SO_2 + 8NH_3 + 8H_2O = 5(NH_4)_2SO_4 \qquad 21)$$

Those processes have been tested with pilot plants (0.2 - 2 MW equivalent) and removed 80 - 90% of NOx with over 95% of SO_2. There are no plans to install larger plants, because the oxidation reduction process requires an expensive oxidizing agent such as O_3 or ClO_2 while the reduction process is complex. Moreover, those processes involve wastewater treatment problems.

5.5 Gas Composition Suitable for Processes

The relationship of SOx and NOx concentrations in the gas to suitable processes is shown in Figure 6. Although the combination of SCR followed by FGD may be used for any composition of gas, the application is not easy for SOx-rich gas because of the ammonium bisulfate problem. Dry simultaneous removal processes may also suit gases with a relatively small SOx/NOx ratio for the following reasons: For the copper oxide process, a high SOx concentration necessitates frequent regeneration which not only requires a large amount of H_2 but also tends to lower NOx removal efficiency. For the carbon process, the carbon consumption increases with SOx concentration. For the electron beam process, the product quality as a fertilizer is low when SOx concentration is high. On the other hand, wet simultaneous removal processes suit SO_2-rich gas because SO_2 works as the reducing agent and increases the NOx removal efficiency.

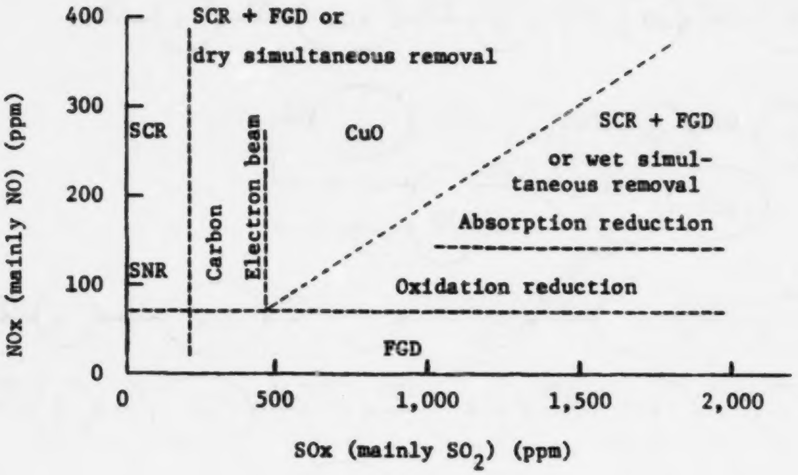

Figure 6 Gas composition and suitable processes

6 ENERGY REQUIREMENT AND COST

6.1 Energy and heat requirement

Figure 7 shows various systems for SO_2 and NOx removal and the power requirement (per cent of power generated by boiler) and heat loss (per cent of heat applied to the boiler) involved in the operation of the systems, with normal boiler operation (No. 1) taken as standard (no requirement, no loss). The energy required for the production of the absorbent, catalyst, and chemicals such as $CaCO_3$, NH_3, H_2, O_3, ClO_2 is not included in the power requirement and heat loss.

No. 2 illustrates an application of FGD (wet lime/limestone process for 90% SO_2 removal) requiring 2-2.5% of power with gas reheating from 55 to 75°C which accounts for 1% heat loss. No. 3 shows an application of SCR to the gas after FGD, which requires 3.0-3.5% power and 5% heat and yet is not free from deposits of ammonium bisulfate as well as solids derived from mists in the heat exchanger and also on the catalyst.

No. 4 is a system that is considered most practical. The gas from a boiler economizer is treated by SCR and then by FGD after heat recovery and dust removal.

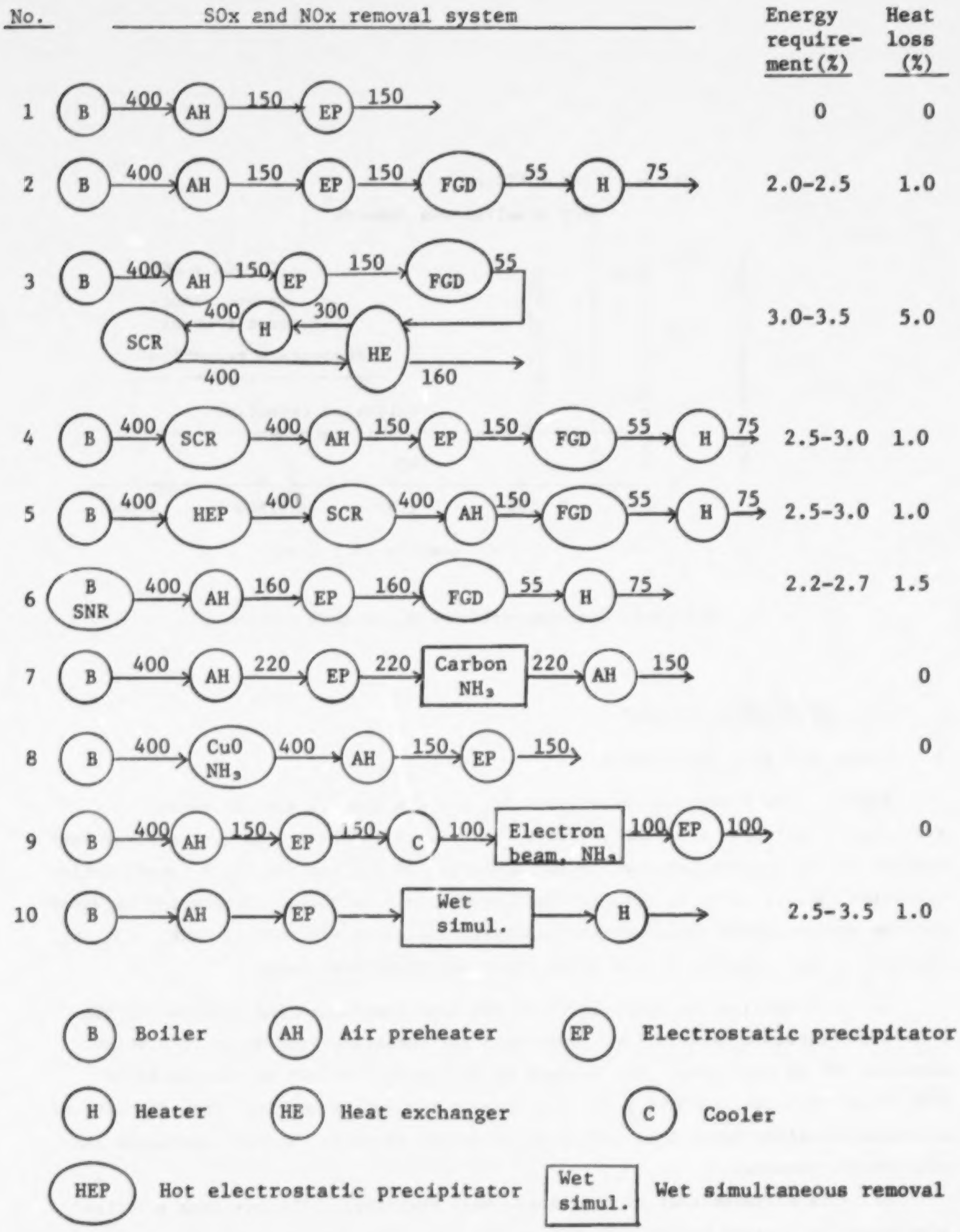
No.
SOx and NOx removal system
Energy requirement (%)
Heat loss (%)
1 B 400 AH 150 EP 150 — 0 0
2 B 400 AH 150 EP 150 FGD 55 H 75 — 2.0-2.5 1.0
3 B 400 AH 150 EP 150 FGD 55 HE 300 H 400 SCR 400 HE 160 — 3.0-3.5 5.0
4 B 400 SCR 400 AH 150 EP 150 FGD 55 H 75 — 2.5-3.0 1.0
5 B 400 HEP 400 SCR 400 AH 150 FGD 55 H 75 — 2.5-3.0 1.0
6 B SNR 400 AH 160 EP 160 FGD 55 H 75 — 2.2-2.7 1.5
7 B 400 AH 220 EP 220 Carbon NH₃ 220 AH 150 — 0
8 B 400 CuO NH₃ 400 AH 150 EP 150 — 0
9 B 400 AH 150 EP 150 C 100 Electron beam, NH₃ 100 EP 100 — 0
10 B AH EP Wet simul. H — 2.5-3.5 1.0
B Boiler
AH Air preheater
EP Electrostatic precipitator
H Heater
HE Heat exchanger
C Cooler
HEP Hot electrostatic precipitator
Wet simul. Wet simultaneous removal

Figure 7 Combined and simultaneous SOx and NOx removal systems
(Figures show temperatures, °C)

Dust removal may be carried out after FGD. One problem with the system is the possible deposit of ammonium bisulfate on SCR catalyst when the gas temperature drops below about 300°C due to the lowering of the boiler load. A temperature drop for a few hours may not hinder the operation because the bisulfate is removed when the gas temperature is raised above 350°C. For temperature drops over longer periods, a heating device such as a hot gas bypass system or an auxiliary burner may be needed. Another problem with the system is the accumulation of ammonia in the wet process FGD system. A device to remove ammonia from the system may be needed.

No. 5 is a system using a hot electrostatic precipitator and may be suitable for flue gas from a low-sulfur coal whose fly ash is not caught efficiently by a cold electrostatic percipitator. As an SCR reactor, not only the parallel flow type but also the moving bed type can be used because the ash content in the gas may be reduced to about 200 mg/Nm^3 by the precipitator. The ammonium bisulfate problem for the air heater may be more serious than with system No. 4 because the gas contains less fly ash, which has a sweeping effect. Other problems are common with No. 4.

No. 6 is a combination of SNR and FGD. Since the NH_3 concentration at the boiler outlet is high, the problems of ammonium bisulfate and ammonia accumulation may be serious. Consequently heat loss may be slightly higher than with Nos. 4 and 5. It may be preferable to place a small amount of SCR catalyst in a duct to reduce leak NH_3 and to increase NOx removal.

Nos. 7-10 show dry and wet simultaneous removal processes. The dry processes require no gas reheating although they require hydrogen, inert gas, or electron beam. Further improvements are desired for commercial application.

6.2 Costs of NOx and SOx Removal

Figure 8 shows investment costs in battery limits of NOx and SOx removal plants with a capacity of up to 1,000,000 Nm^3/hr of flue gas from a boiler, which is equivalent to 330 MW with oil and 280 MW with coal. Figure 9 shows annualized operation costs for the plant at 8,000 hours' yearly operation including 7 years' depreciation assuming the total investment cost is 50% more than the investment cost in battery limits. The costs are based on investigations by the Japan Environment Agency[11)] and modified by Ando based on his study. The cost per kW or kWhr for flue gas from coal is estimated at 20% more than that for flue gas from oil because of the larger gas volume per kW.

The cost for SCR by the direct process, viz., treating directly the flue gas from a boiler economizer, are 3,000-4,000 yen/kW for investment and 0.3-0.4 yen/kWhr

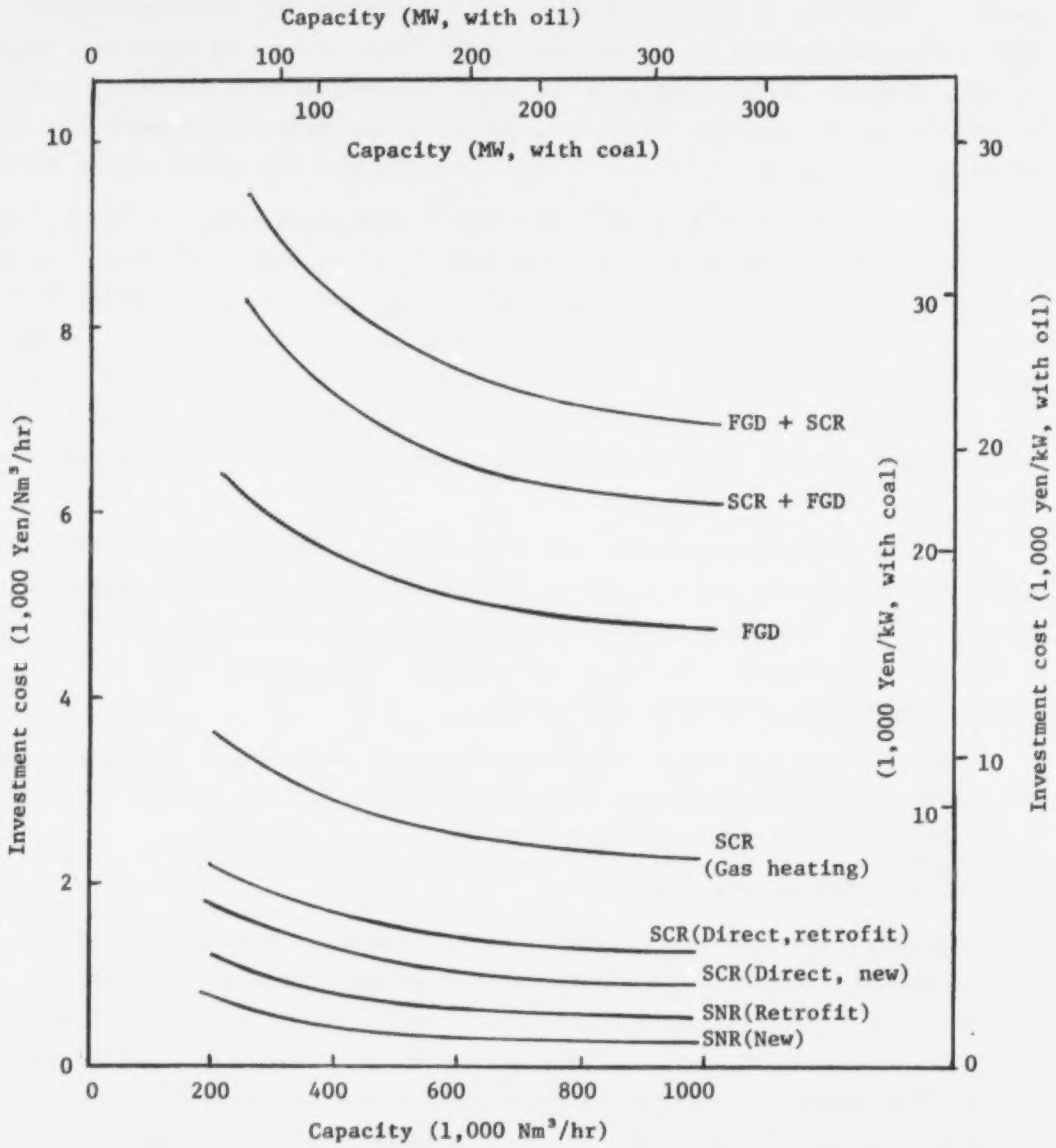

Figure 8 Investment in battery limits of NOx and SOx removal plants

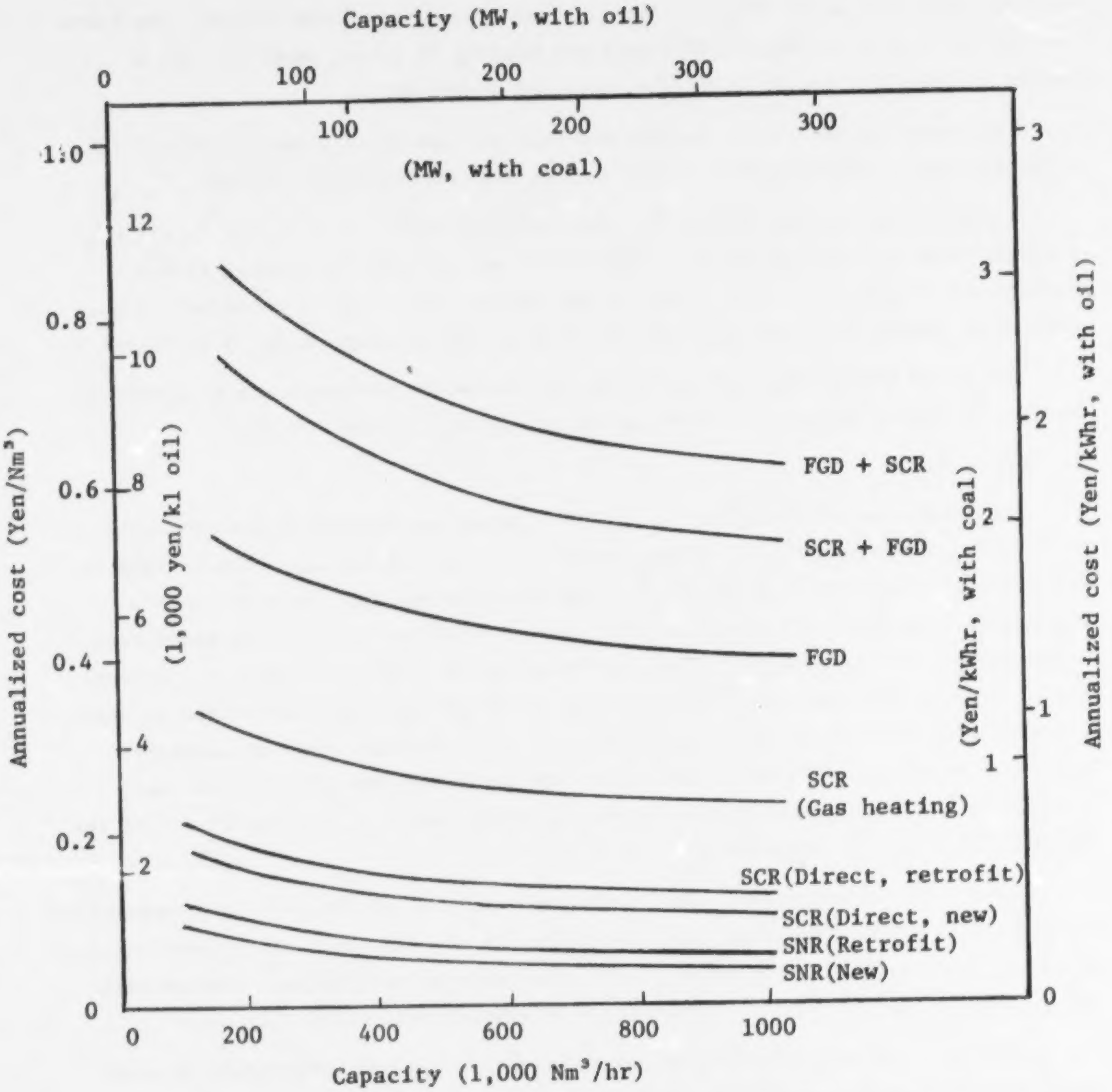

Figure 9 Annualized operation cost for NOx and SOx removal plants

for operation with a 200 MW plant for flue gas from an oil-fired boiler. The costs for SCR for a cold gas about 150°C with gas heating by a heat exchanger and a heater to 300-400°C are about double those for direct SCR.

The costs for SNR may be between one-half and one-third those for direct SCR, while NOx removal efficiency is about half in SCR (40-45% versus 80-90%).

A combination of SCR followed by FGD, as shown in No. 4 of Figure 7, may cost a little more than the sum of the costs for direct SCR and FGD because of the requirement of ammonia removal from the FGD system, but is still considerably more economical than a combination of FGD followed by SCR as shown in No. 3 of Figure 7.

The costs for the dry and wet simultaneous removal processes are uncertain but seem to be higher than those for a combination of SCR followed by FGD.

7. CONCLUSION

The recent remarkable progress of FGD in Japan was induced by the following particular circumstances: (1) Severe public criticism and stringent regulations on pollution. (2) Control by governments using telemeter systems. (3) Government's assistance to industry by providing low-interest funds and by allowing short-term depreciation. (4) Most of the plants were constructed while the Japanese industry was growing rapidly; the total investment for FGD and hydrodesulfurization of heavy oil amounting to 1 trillion yen at the current value did not prove an excessive burden on industry. (5) Lime scrubbing and ammonia scrubbing which were already applied in the 1950s provided a basis for the development. (6) Virtually all of the by-products have been utilized.

Under such circumstances, over 1,000 commercial FGD plants have been constructed, operated with good performance, and contributed to the abatement of SO_2 concentration. Although the desulfurization efforts have attained the goal, further studies will be desired to improve FGD --- particularly in simplification and cost cuts.

Concerning NOx removal, combustion modification has made remarkable progress. Moreover, SCR has been improved considerably and proved commercially applicable to flue gases from gas and low-sulfur oil burning. Further improvements are expected to apply SCR to more dirty gases. SNR may be suitable for certain gas sources. Combination of SNR and SCR may also be useful. For simultaneous SOx and NOx removal processes, further tests are needed to evaluate the feasibility of commercial application.

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4E

EPRI'S FGD PROGRAM: FROM PROBLEM IDENTIFICATION TO DEVELOPMENT OF SOLUTIONS

G. T. Preston
Electric Power Research Institute
Palo Alto, CA 94303

ABSTRACT

The overall objective of the Desulfurization Processes Program at Electric Power Research Institute (EPRI) is to develop the most reliable and cost effective flue gas desulfurization (FGD) technologies that can satisfy regulatory requirements. Through 1978 this objective was pursued primarily through evaluation of established techniques and development of emerging processes. Representative projects reviewed in this paper are lime scrubbing and sludge disposal guidelines; characterization of full-scale operating utility FGD systems; and pilot and laboratory efforts in absorption/steam stripping, RESOX, and an aqueous carbonate sodium regeneration process. In 1979, the emphasis of EPRI's program is shifting to evaluation and demonstration of advanced technologies, ranging from prototype evaluation of the cocurrent scrubber configuration and the Chiyoda Thoroughbred 121 and RESOX processes, to the start of construction of one or more full-scale (100 MW) FGD systems.

EPRI'S FGD PROGRAM: FROM PROBLEM IDENTIFICATION TO DEVELOPMENT OF SOLUTIONS

The purpose of this paper is to provide an overview of the flue gas desulfurization research program at Electric Power Research Institute. The structure and goals of the program are described briefly and the program staff are introduced, followed by a summary of significant results achieved in each of four subprogram areas within the program, since our last report to this symposium.[1]

PROGRAM STRUCTURE

Electric Power Research Institute (EPRI) is the research and development arm of the U.S. electric utility industry. It is a not-for-profit organization funded by membership assessments from public and private electric utilities representing about 80% of U.S. generating capacity. EPRI promotes the development of new and improved technologies to help the utility industry meet present and future electric energy needs in environmentally acceptable ways.

Responsibility for research in flue gas desulfurization by stack gas scrubbing rests with the Desulfurization Processes Program, part of EPRI's Fossil Fuel Power Plants Department. The overall objective of the Program is to develop the most cost-effective flue gas desulfurization (FGD) technologies that can satisfy regulatory requirements. The abatement of sulfur oxide emissions from utility boilers is presently a technology-forced challenge; that is, clean air regulations currently being promulgated are based on the most advanced existing technologies, even though some of them have been demonstrated at only relatively small scale.[2,3] This regulatory approach exerts pressure on EPRI to identify the FGD technologies that are fundamentally sound and to move these toward commercial feasibility as rapidly as possible. Accordingly, the three-phase strategy of the Desulfurization Processes Program is to: evaluate on a continuing basis the established technologies and develop design and operating guidelines; evaluate and develop emerging technologies through pilot plant tests and operation of integrated prototype systems; and participate in the demonstration of commercial-scale scrubbing systems.

The four subprograms of EPRI's FGD program, and the objectives of each subprogram over the next five years, are shown in Table 1. Activities in each subprogram follow the three-phase strategy of evaluation, pilot and prototype testing, and demonstration at commercial scale.

As EPRI's FGD program has grown rapidly, the breadth of identified research needs has outstripped the funding and staff available to address those needs. This forces a careful evaluation of each potential research topic to insure that we make the most effective use of our resources. Does it address the overall objective of our program? Does it fit our established strategy? And, does it address one or more of our subprogram objectives?

There is a further requirement that we impose: that a potential research project address one, or preferably several, of the specific technical and economic incentives associated with doing research in flue gas desulfurization. These are as follows:

Technical: Compliance with SO_2 removal standards

Maximum system reliability

Minimum water consumption

Minimum by-product volume, or maximum by-product value

Minimum heat rate penalty (energy consumption)

Economic: Minimum capital cost

Minimum operating and maintenance cost

AVAILABILITY OF RESULTS

One of EPRI's charges as a tax-exempt organization is to make the results of the research available for the public benefit. This is done in a formal way by publishing an EPRI report at the end of each research project. A copy of the EPRI

Table 1. EPRI'S DESULFURIZATION PROCESSES PROGRAM

SUBPROGRAM	OBJECTIVES (1979-1983)
State of the Art	Maintain up-to-date design and operating guidelines for alkali scrubbing Characterize performance, reliability and cost of full-scale utility FGD systems Establish scrubber operator training centers
Subsystem Evaluation and Development	Evaluate and develop energy-conserving scrubber subsystems Compile materials of construction experience Develop advanced contactors and scrubber configurations
Advanced FGD With Recovery	Demonstrate advanced FGD to recover sulfur using coal reductant Demonstrate sodium regeneration subsystem in a sulfur-recovery process
Advanced FGD Without Recovery	Demonstrate sludge-free lime/limestone FGD processes Develop improved chemistry basis for alkali scrubbing

Publications List may be obtained from Research Reports Center, Post Office Box 10090, Palo Alto, CA 94303, (415) 961-9043. Reports which have been published in the FGD area are listed in Table 2.

Information on research in the Desulfurization Processes Program in particular is available on an informal basis through direct contact with the program manager or one of the project managers; they are identified in Table 3.

STATE OF THE ART

The overall goal of this subprogram is accurately and on a timely basis to assess the status of FGD technology as it is commercially available to the utilities, so that a utility with a need to install SO_2 scrubbing has up to date information on which to base procurement decisions, and so that operators of existing FGD systems can maximize reliability and minimize their cost. Recent results in this area are summarized below.

Stack Gas Emission Coordination Control Center

The objective is to maintain an up-to-date information base of scrubber operating experience and data as an aid to EPRI member utilities in keeping their FGD systems operating and in planning future installations, and to identify common operating problems as an aid to EPRI research planning. Battelle Columbus Laboratories is the contractor.

Since our last report to this symposium, Battelle has completed an analysis of the causes of wide variations in the bid-price cost estimates for FGD systems. The principal factors are the date of the estimate and the estimating procedure. The impact of the estimate date is due not only to inflation but also to design changes resulting from technological advances and new regulatory requirements.

As part of this project, Battelle provides short term information search and consulting efforts without charge to EPRI member utilities who have specific information needs covering any aspect of FGD technology. Details can be obtained from Dr. Harvey Rosenberg at Battelle, (644) 424-5010, or from Tom Morasky at EPRI.

Table 2. PUBLISHED REPORTS - DESULFURIZATION PROCESSES PROGRAM

DATE	REPORT NO.	TITLE
Aug 1975	EPRI 209 Part 1	Status of Stack Gas Control Technology
Aug 1975	EPRI 209 Part 2	Status of Stack Gas Technology for SO_2 Control
Sep 1975	EPRI 202	Environmental Effects of Trace Elements from Ponded Ash and Scrubber Sludge
Mar 1976	FP-272 Vol. I,II Addendum	Evaluation of Regenerable Flue Gas Desulfurization Processes
Oct 1976	FP-207	Evaluation of Dry Alkalis for Removing Sulfur Dioxide from Boiler Flue Gases
Dec 1976	FP327	Guidelines for the Design of Mist Eliminators for Lime/Limestone Scrubbing Systems
Feb 1977	FP-361	Stack Gas Reheat for Wet Flue Gas Desulfurization Systems
Jul 1977	FP-463-SR	Process Synthesis and Innovation in Flue Gas Desulfurization
Dec 1977	FP-595	Application of Scrubbing Systems to Low Sulfur Alkaline Ash Coals
Dec 1977	FP-639	A Summary of the Effects of Important Chemical Variables Upon the Performance of Lime/ Limestone Wet Scrubbing Systems
Mar 1978	FP-713 Vol. I-III	Evaluation of Three 20 MW Prototype Flue Gas Desulfurization Processes
Mar 1978	FP-671 Vol. III	State-of-the-Art of FGD Sludge Fixation
Mar 1978	FP-889	EPRI/Radian Particle Balance Concept Study
Oct 1978	FP-909	Analysis of Variations in Costs of FGD Systems
Jan 1979	FP-942	Full-Scale Scrubber Sludge Characterization Studies

Table 2. PUBLISHED REPORTS - DESULFURIZATION PROCESSES PROGRAM (Continued)

DATE	REPORT NO.	TITLE
Jan 1979	FP-671 Vol. 1	Review and Assessment of the Existing Data Base Regarding Flue Gas Cleaning Wastes
Jan 1979	FP-977	FGD Sludge Disposal Manual
Jan 1979	FP-941	Cocurrent Scrubber Evaluation TVA's Colbert Lime/Limestone Wet Scrubbing Pilot Plant
Feb 1979	FP-671 Vol. II	Chemical/Physical Stability of Flue Gas Cleaning Wastes
Mar 1979	FP-1030	Lime FGD Systems Data Book

Table 3. EPRI FGD STAFF

George T. Preston (415) 855-2461	Program Manager
Stuart M. Dalton (415) 855-2467	Project Manager State of the Art Advanced FGD with Recovery
Charles E. Dene (502) 443-6489	Facility Manager - Shawnee State of the Art
Thomas M. Morasky (415) 855-2468	Project Manager State of the Art Advanced FGD without Recovery
Richard G. Rhudy (415) 855-2421	Project Manager State of the Art Subsystems Evaluation & Development Advanced FGD with Recovery
Dorothy A. Stewart (415) 855-2609	Project Manager Subsystems Evaluation & Development Advanced FGD without Recovery

Evaluation of Improved Process Control Capability for FGD Systems

The overall objective is to evaluate the status of FGD process control practices and instrumentation and recommend changes and/or research to develop improved methods. The project has been completed since our last report to this symposium, and three reports have been published. The major conclusions are:

- By-product sludge characteristics can be improved by operating FGD systems so as to reduce the rate of formation of new sludge crystals.

- Poor operating reliability in FGD systems is often due to a spiraling sequence of: scale formation, sensor malfunction, unstable control of pH and other chemical concentrations, and a further increased rate of scale formation. Valid sampling procedures are essential in maintaining effective operating control.

Follow-on research is planned for late 1979 or 1980, to develop improved instrumentation and sampling techniques to break the instability/scaling cycle referred to above.

Characterization of Full-Scale Scrubbers

The objective is to characterize and publish data for four representative full-scale utility lime and limestone wet scrubbing systems. Field testing is being carried out to determine removal efficiencies for regulated air and water emissions, such as SO_2 and particulates, as well as currently unregulated discharges such as fine particulate, polycyclic organics, and vapor-phase metals. An engineering/economic evaluation of each system is also being performed to document system operability and costs. The four systems currently scheduled for characterization are Pennsylvania Power Company's Bruce Mansfield plant, Columbus & Southern Ohio's Conesville plant, Montana Power's Colstrip station, and Northern States Power's Sherburne County station. These four units constitute a cross-section of lime and limestone FGD technology on western and eastern coal. The first phase of testing has been completed at the Conesville plant, and testing will begin at Colstrip in April 1979.

By-Product/Waste Disposal for Flue Gas Cleaning Processes

The overall objective of several research efforts under this project is to establish a sound information base to help utilities select fly ash and scrubber sludge disposal methods from commercially available technologies. The project is complete; four reports have been issued describing the following results:

- The existing information base on flue gas cleaning (FGC) wastes has been reviewed and a laboratory program was outlined to fill in major gaps in the available knowledge.

- The laboratory program is continuing. One useful preliminary conclusion is that it is possible to predict the long-term stability of fly-ash-stabilized sludge with reasonable confidence on the basis of its physical characteristics after 50 days aging.

- The FGD sludge fixation processes of IUCS and Dravo are sufficiently developed and tested to be considered commercially available for utility application. The incremental costs of sludge fixation over disposal by ponding or landfilling are estimated at \$6.90 and \$2.50, respectively, per ton of dry sludge for a hypothetical 1000 MW generating station. (Since these estimates reflect only the additional cost of fixation, they do not indicate whether ponding or landfill will be preferred at a given site.)

- Sludge slurry samples from six operating full-scale utility FGD systems were evaluated for their dewatering characteristics when subjected to settling, filtration, or centrifugation. The results confirmed that larger sludge particle size distributions lead to better dewatering characteristics, and indicated also that the use of flocculating agents can usually improve dewatering performance and reduce cost. It appears that quantitative understanding of the relationship between dewatering characteristics and scrubber operating parameters cannot be derived from data obtainable in a full-scale system in normal day-to-day operation.

- A comprehensive guidelines manual for disposal of FGD sludges has been published which provides detailed information on available technology; design, specification, and procurement; and operation of FGD sludge disposal systems. The manual incorporates the results of the earlier EPRI work described above.

Lime FGD Systems Data Book

The objective is to compile lime scrubbing design and operating guidelines for utility FGD applications. The project is complete, and the final report will be available by mid-April. All the FGD aspects investigated in the State of the Art and the Subsystems Evaluation and Development subprograms over the past two years are included in the manual as they apply to lime scrubbing systems; additional topics included are Process Design, Equipment Design, Procurement Procedures, Lime Handling, Slurry Preparation and Corrosion. Shortly after publication of the report EPRI will sponsor a workshop to familiarize utility staff with the contents of the guidelines and to elaborate on how they can be used most effectively.

SUBSYSTEMS EVALUATION AND DEVELOPMENT

The overall goal of this subprogram is to identify problems and develop solutions in areas of technology which are not directed toward a single FGD process but are related to wet scrubbing systems in general; such areas include mist elimination, reheat, materials of construction, and novel gas/liquid contacting configurations. Recent important progress in this subprogram is summarized below.

Improved Lime/Limestone Scrubbing Technology

The objectives were to evaluate two innovative gas/liquid contactor configurations, to investigate the concept of stack gas reheat, to correlate sludge properties with scrubber design and operating conditions, and to establish an information base on corrosion and erosion to aid in materials of construction selection. The work was carried out by TVA, including 1 MW testing at the Colbert Steam Plant. Reports on all these subjects have been issued or are in press, and most of these efforts have already led to larger-scope follow-on research. The major conclusions reached are as follows:

- SO_2 removal efficiencies in a pilot-scale horizontal (cross-flow) scrubber were primarily affected by liquid rate, gas velocity, pressure drop across the slurry nozzles, and gas/droplet contact time. In general, the removal efficiencies were lower than had been expected, perhaps because the gas/droplet contact time achievable in a commercial scale module could not be simulated in the pilot unit. Even so, under certain test conditions, SO_2 removal efficiencies of over 90% were obtained with each of the two reagents (lime and limestone) tested.

- SO_2 removal efficiencies in a pilot scale cocurrent flow scrubber were sensitive to where the scrubbing liquor was introduced into the scrubber and the presence or absence of open (grid) packing in the scrubber to improve gas and liquid distribution and increase the gas/liquid contact time. Liquid rate generally was a more significant variable than gas velocity in affecting SO_2 removals. These tests provided the basis for design of a 10 MW cocurrent scrubber at TVA's Shawnee Test Facility.

- Laboratory testing of the physical and chemical properties of sludges from operating utility FGD systems showed that the physical form in which calcium sulfite precipitates in a scrubber system depends on whether the source of calcium is lime or limestone. Calcium sulfite sludge from a limestone FGD system consists of simple, open structure, tabular crystals. In contrast, calcium sulfite solids from lime FGD systems are complex, interpenetrating, spheroidal aggregates. This study showed that the settled bulk density of FGD sludges decreases with increasing solids surface area, meaning that calcium sulfite sludges from lime FGD systems dewater much less readily than those from limestone systems.

- Capital costs for a 1 MW flue gas recirculation reheat system were about 75% higher than for an in-line indirect steam reheater, and the estimated operating cost was about 9% higher. However, elimination of several potential operating problems inherent in an in-line system, such as plugging, corrosion, and pitting of the shell side of the reheater tubes due to moisture carryover from the scrubber, might justify the added cost for the flue gas recirculation reheat approach. Careful attention to reheater operating conditions and effective mist elimination can minimize the potential problems with in-line reheat.

In conjunction with the in-line indirect steam reheater evaluation, several materials of construction were tested for resistance to erosion and corrosion. Type 316 stainless steel and Incoloy 825 showed very good resistance. Incoloy 800; types 304, 410, and 446 stainless; U.S. Steel alloy 100; and 18-18-2 alloy showed fair to good resistance to erosion, but suffered from pitting and crevice corrosion. Cor-Ten alloys A and B showed high rates of erosion/corrosion.

Advanced Flue Gas Desulfurization Development and Test Facility

The overall objective is to construct and operate 10 MW prototype scrubber facilities to evaluate advanced scrubber concepts. This is accomplished at TVA's Shawnee Test Facility; EPRI currently supports the operation of one of three 10 MW prototype FGD units, a cocurrent scrubber system which was substantially completed in September 1978. Testing has been completed using sodium carbonate, magnesium oxide, lime and limestone as reagents. Reliability testing with limestone is in progress.

In the latter half of 1979, another of the Shawnee 10 MW systems will be used to evaluate the Dowa Mining process; this is described below under the Advanced FGD Without Recovery subprogram. In 1980 EPRI hopes to see one or more of the Shawnee prototype systems in use as a utility staff training center and a debugging tool to solve FGD operating problems at full-scale units. That effective use of the Shawnee facility for such a purpose could improve the reliability of existing systems drastically is suggested by the strong historical correlation between high FGD reliability and the presence of well-trained operating and maintenance staff.

Other Projects

Additional efforts in progress or being initiated in the Subsystems Evaluation and Development subprogram include:

- Construction Materials for Wet Scrubbers
- Scrubber Generated Particulates
- Cyclic Reheat
- Entrainment in Wet Stacks

ADVANCED FGD WITH RECOVERY

The overall goal of this subprogram is to encourage the commercialization of advanced SO_2 control techniques which involve recovery of the sulfur in a marketable form such as elemental sulfur or sulfuric acid. The two major FGD approaches EPRI is currently supporting in this area are absorption/steam stripping/RESOX and the Aqueous Carbonate Process.

Absorption/Steam Stripping/RESOX tm

Several projects are directed toward the objective of demonstrating an advanced FGD process featuring sulfur recovery without the need for a gaseous reductant. The combination of absorption/steam stripping with RESOX has the potential to eliminate FGD sludge disposal problems and produce elemental sulfur at a cost comparable to conventional lime/limestone scrubbing. Sulfur oxides are absorbed in a buffer solution and then steam stripped from the solution in a separate vessel. Thus the effect of absorption/steam stripping is to concentrate SO_2 from several thousand ppm to 25-95% by volume, the balance being water vapor. In RESOX, the concentrated SO_2 stream is reduced to elemental sulfur by contact with a bed of crushed coal. Absorption/steam strip chemistries are at various stages of development by about a half-dozen suppliers, while RESOX is a proprietary process of Foster Wheeler Energy Corporation. This particular combination of the two subsystems is not unique--that is, each of them might be used in other combinations. However, absorption/steam strip/RESOX appears to be among the lowest cost sulfur recovery FGD options. Recent results achieved in EPRI research efforts are as follows.

- Laboratory-scale tests were carried out to establish vapor/liquid equilibria for three absorption/steam stripping chemistries. The Flakt-Boliden sodium citrate buffer process was selected for pilot testing.

- Preliminary engineering is complete and long lead time equipment items have been ordered for a 1 MW absorption/steam stripping pilot test program at TVA's Colbert Steam Plant.

- A 42 MW prototype RESOX plant was constructed and started up on schedule in West Germany. The concentrated SO_2 feed stream was converted to elemental sulfur of over 99% purity, using German anthracite coal as the reductant. One purpose of this work is to evaluate the applicability of RESOX to several SO_2-concentrating front-end subsystems--Bergbau-Forschung activated carbon, absorption/steam stripping, magnesia and Wellman Lord.

- Two U.S. noncaking bituminous coals were shown in lab testing to be suitable for use as the RESOX reductant. Caking bituminous coals are not suitable unless they have been pretreated to eliminate agglomeration characteristics.

Aqueous Carbonate Process

Two EPRI projects address the objective of demonstrating a sodium regeneration subsystem applicable to sulfur recovery FGD. Both are directed toward commercialization of Rockwell International's Aqueous Carbonate Process (ACP). Sulfur oxides are absorbed in an aqueous solution of sodium carbonate. The contacting device is a spray dryer. The dry particles of spent absorbent are collected by a baghouse or an electrostatic precipitator and charged to a molten salt reducer which uses coal to convert sodium sulfite and sulfate to sulfide. The reducer and the subsequent carbonation system which regenerates the absorbent sodium carbonate solution potentially can be applied to other sulfur-recovery FGD processes such as Wellman Lord and magnesium oxide scrubbing.

EPRI has expressed its intention, subject to Board of Directors approval, to participate in a 100 MW demonstration of ACP at the Huntley Station of Niagara Mohawk Power Corporation. Empire State Electric Energy Research Corporation is the lead agency for this demonstration; EPA and New York State ERDA are also participants. Preliminary engineering for the demonstration is proceeding now.

One of the EPRI projects, cofunded by Niagara Mohawk, has the objective of obtaining test data on a 5 MW pilot scale for use in the design of the 100 MW plant. The other project, preparation of a test requirements document, has the objective of assuring that the technical requirements of all the demonstration project participants are addressed in the final design of the 100 MW plant.

ADVANCED FGD WITHOUT RECOVERY

The overall goal of this subprogram is to develop and demonstrate scrubbing processes which offer significant improvement over conventional FGD technology in the areas of removal efficiency, reliability, sludge disposal, and cost.

Sludge-Free Limestone Scrubbing

The Chiyoda Thoroughbred 121 Process accomplishes limestone dissolution, SO_2 absorption, sulfite oxidation to sulfate, and by product gypsum thickening, all in one flue gas sparged reactor vessel. The objectives of the project are:

- Evaluate the performance, control characteristics, operating flexibility and reliability of the CT-121 process at a prototype scale.
- Determine the feasibility of disposing of forced oxidation gypsum by-product solids by stacking.

The results achieved to date are as follows.

- The 23 MW CT-121 prototype facility was constructed and started up (at Chiyoda expense) at Gulf Power Company's Plant Scholz.
- Laboratory tests established the feasibility of stacking the CT-121 by-product gypsum as a disposal technique.
- A 5 month evaluation of the CT-121 process is in progress. SO_2 removals have usually been over 90%.

- The gypsum stack has been established, and monitoring of its strength, permeability, and leachate composition is in progress. So far, the physical characteristics of the stack confirm the feasibility of stacking the CT-121 by-product.

EPRI and TVA are planning the evaluation of another sludge-free limestone-based advanced process, the Dowa Mining Dual Alkali Process. SO_2 is absorbed in a basic aluminum sulfate solution. Limestone is used to regenerate the aluminum values, and forced oxidation results again in rejection of sulfur as a gypsum by-product. In addition to the potential benefits from clear solution scrubbing and production of gypsum to decrease disposal costs, a significant attraction of the Dowa process is its potential for retrofit to existing lime or limestone scrubber systems. The Dowa evaluation will be at one of the 10 MW prototype scrubber systems at TVA's Shawnee Test Facility. EPRI anticipates that results from this project will be available in the spring of 1980.

Chemical Basis of Alkali Scrubbing

It is obvious that the performance of FGD systems generally, and SO_2 removal efficiency and system reliability in particular, are directly related to the chemistry of the scrubbing process. Shortly after our last report to this symposium, EPRI initiated a project whose objective was to quantify several chemical phenomena in lime and limestone scrubbing through the development of mathematical models from fundamental chemistry and mass transfer principles, and fitting of these models to the available data. Six aspects of lime and limestone FGD chemistry were the subjects of modelling efforts: SO_2 removal efficiency, oxidation, gypsum supersaturation, gypsum subsaturation, solids quality, and alkali utilization. This project is nearing completion. A conclusion which is already apparent is that sufficiently reliable and internally consistent data on which to base the models are much scarcer than we had thought originally. The implication of this is that although substantial effort has been directed to data collection in prototype and full-scale scrubber systems, those data were not taken under conditions or in a manner such that they are useful in arriving at a fundamental understanding of the chemical phenomena of scrubbing. Rather than pilot plant testing of the models, then, the next step in the EPRI effort will likely be bench-scale studies to fill in the gaps that have been identified as a result of this work.

Other Projects

Other efforts in progress or being initiated in the Advanced FGD Without Recovery subprogram include:

- Design of High SO_2 Removal FGD Processes
- Cost of High SO_2 Removal FGD Processes
- Spray Drying FGD Evaluations
- FGD Reagent Preparation

FULL-SCALE DEMONSTRATIONS

As a result of earlier evaluation studies and some of the pilot-scale efforts described above, four advanced FGD processes have been selected for potential major EPRI funding and participation through the 100 MW demonstration stage. The processes are Chiyoda Thoroughbred 121, Dowa Mining, absorption/steam strip/RESOX, and Aqueous Carbonate. The planned pilot-scale and prototype evaluations will be complete for all except absorption/steam stripping and Dowa by mid-1979. Therefore, EPRI is now seeking utility sites for the full-scale demonstrations.

CONCLUSION

Since our last report to this symposium, EPRI's flue gas desulfurization research effort has evolved from subprogram to full program status; has tripled in staff, budget, and number of active projects; and has shifted from an initial mode of surveying the state of the art and identifying problems, to one of developing, testing, evaluating, and demonstrating solutions to those problems.

ACKNOWLEDGMENTS

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Niagara Mohawk Power Corp.
Steag AG
Tennessee Valley Authority
UOP, Inc.

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Title

CHOLLA STATION UNIT 1 FGD SYSTEM 5 YEARS OF OPERATING EXPERIENCE

Authors

Stephen R. Travis
Chemical Engineer, Technical Services
Electric Operations
Arizona Public Service Company

Frank. A. Heacock, Jr.
Manager, Mechanical Technical Services
Arizona Public Service Company

Abstract

The Cholla Unit 1 FGD limestone throwaway system has been in commercial operation since December 14, 1973. The operations of this system has been characterized by high efficiency control of flue gas SO_2 and particulate matter at a high, sustained reliability factor. Trends in O&M costs demonstrate sound initial engineering and O&M concepts. Five years of operation give a data base of useful information on system component reliability and maintenance trends. Data available for performance of materials of construction in severe service is presented to support the use of corrosion resistant alloy materials.

Based on the data and trends presented, extrapolations are discussed for process capabilities and for design and selection of equipment by generic type for reliable operation and acceptable maintenance.

CHOLLA STATION UNIT 1 FGD SYSTEM
5 YEARS OF OPERATING EXPERIENCE

INTRODUCTION

The FGD system for Unit 1 at Cholla Station of Arizona Public Service marked five years of continuous service in December, 1978. For the entire period, this double-loop, limestone, throwaway FGD system has operated at a high reliability factor and with high efficiencies both in required flue gas SO_2 removal and particulate removal.

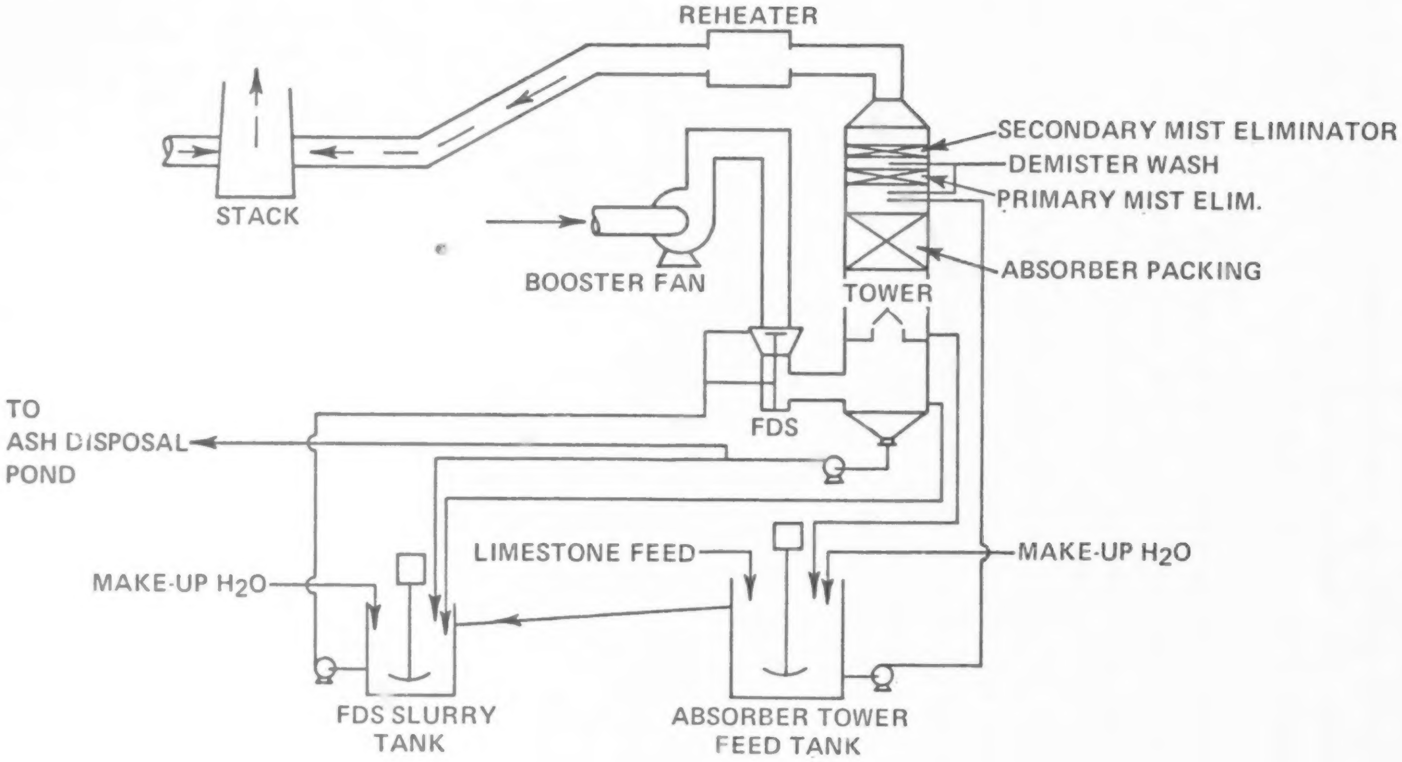
The FGD system for this 115 MW generating unit consists of two parallel gas cleaning trains. Each gas cleaning train consists of a high energy Flooded Disc type venturi scrubber, and a wetted-film contact absorber.

Detailed descriptions of the FGD system components and process have been presented previously by L. K. Mundth, 1974, (1), PEDCo Report 1978 (2), and in various trade journal articles. For purposes of this paper, only a brief description will be given. Reference is made to Figure 1 for a schematic presentation of the components and process.

The FGD system was a retrofit to the draft system of the unit and, therefore, takes suction from the discharge of the existing ID fans. These fans are preceded by multicyclone mechanical dust collectors. The existing breeching to the unit chimney forms a full bypass alternate for the FGD system.

As Figure 1 indicates, flue gas flows thru the FDS treatment loop for removal of particulate materials and partial removal of SO_2 gases. The flue gas then passes vertically upward thru the combined cyclonic entrainment separator/wetted-film absorber tower, thru the two stage mist eliminator, vertically downward thru the reheater and exhausts into the chimney.

Figure 1. CHOLLA UNIT 1 WET SCRUBBER SYSTEM


REHEATER
STACK
SECONDARY MIST ELIMINATOR
DEMISTER WASH
PRIMARY MIST ELIM.
BOOSTER FAN
ABSORBER PACKING
TOWER
FDS
TO
ASH DISPOSAL
POND
LIMESTONE FEED
MAKE-UP H2O
MAKE-UP H2O
FDS SLURRY
TANK
ABSORBER TOWER
FEED TANK

Approximately 25% of the flue gas SO_2 is absorbed in the FDS loop in each train along with 99+% of the particulate matter. Only train A is fitted with the wetted-film absorber packing and, in this train, the combined overall removal of SO_2 is designed for 92%.

Table 1 contains some pertinent data as reported in Table 4 of the PEDCo Report, 1978 (2) (by Permission). The data relates to hold tank design but is at least introductory for purposes of this paper.

The flyash/sludge wastes are currently bled from the FDS loop and pumped for ultimate disposal in a new evaporative disposal area for the station. The FGD system operates on an open-water-loop basis with no recycle of water from the pond. Fresh water make-up to the FGD system comes from the unit cooling water lake or from deep wells.

The sustained success of this FGD application is in great part dependent upon the unique factors of waste disposal and open-water-loop operation and underwrites the argument that FGD system applications must be site specific and that generalizations of successful experience at one site may not be viable for another.

Table 1. DATA SUMMARY: FGD SYSTEM HOLD TANKS

	Flooded disc scrubber hold-up tank	SO_2 absorber towers hold-up tank	FGD system sludge hold-up tank	Limestone slurry make-up tank
Total number of tanks	One	One (common)	Two	Two
Tank sizes	3.8 m (12.5 ft) dia. x 4.3 m (14 ft)	8.3 m (27.3 ft) dia. x 8.5 m (28 ft)	5.6 m (18.5 ft) dia. x 8.2 (27 ft)	
Retention time at full load	7 min	5 min	14 hr	
Temperature	49°C (121°F)	49°C (121°F)	49°C (121°F)	32°C (90°F)
pH	5.2	6.5	5.2	
Solids concentration, percent	15.5	8.3	25	20
Specific gravity	1.102	1.049		

PERFORMANCE

Initial FGD system performance testing was conducted in October, 1973. Table 2 contains the operational data and performance results of this testing. This is as reported by the PEDCo report, 1978 (2) (by permission).

As shown in Table 2, the SO_2 removal efficiency during the initial testing indicated 92.4 percent in the A-side and 14.4 percent for one of two test sets in the B-side for a combined average of 53.4 percent. Also, the particulate removal efficiency was 99.7 percent in the A-side and 99.8 percent in the B-side.

In October 1977, further testing at the FGD outlet was conducted which indicated a combined average SO_2 removal efficiency of 43.0 percent and a combined particulate removal efficiency of 99.75 percent. For these October 1977 tests, inlet loadings were estimated from coal date.

OPERATION

Reliability

The reliability of the FGD system has been consistently high as shown in Table 3. During the period from January, 1974 to November 1978, the reliability of the A-side averaged 93.0 percent while the B-side averaged 91.4 percent for a combined average of 92.2 percent. In this context, reliability is defined as the hours the FGD system was operated divided by the hours the FGD system was called upon to operate.

Table 2. RESULTS OF FGD SYSTEM PERFORMANCE TEST RUNS, OCTOBER 2 to 21, 1973

	A-side	B-side	B-side
Particulate concentration inlet, g/m^3 (gr/scfd)	4.569 (1.995)	5.810 (2.537)	
Particulate concentration outlet, gr/scfd	0.0190 (0.0083)	0.0231 (0.0101)	0.2631 (0.1149)
SO$_2$ concentration outlet, ppm	34	357	236
SO$_2$ concentration inlet	417	409	
Configuration	Packed	Hollow	Hollow
SO$_2$ removal, percent	92.4	14.4	9.2
Particulate removal efficiency, percent	99.7	99.8	
Gas inlet to FDS. m^3/sec (acfm)	96.9 (214,300)	96.6 (204,600)	96.4 (204,300)
Theoretical inlet gas to FDS, m^3/sec (acfm)	93.8 (198,000)	93.8 (198,800)	93.8 (198,800)
Apparent bypass leakage, m^3/sec (acfm)		7.98 (16,900)	
FDS L/G ratio, liters/m^3 (gal./1000 acf)	1.35 (10.1)	1.42 (10.6)	0.78 (5.8)
Tower L/G ratio, liters/m^3 (gal./1000 acf)	6.5 (48.9)		

(continued)

Table 2 (continued)

	A-side	B-side	B-side
Gas velocity through tower, m/sec (ft/sec)	2.10 (6.9)	2.05 (6.6)	2.05 (6.6)
Mist entrainment from tower g/m^3 (gr/scf)	0.000	0.000	NA
Solids entrainment from tower slurry g/m^3 (gr/scf)	0.011 (0.005)	NA	NA
Pressure drop FDS, kPA (in. H_2O)	3.7 (14.8)	3.9 (15.7)	3.9 (15.7)
Pressure drop tower demisters, kPA (in H_2O)	0.0	0.0	
Pressure drop reheater, kPa (in. H_2O)	1.3 (5.15)	0.6 (2.30)	
NA -Not applicable.			
Temperature tower outlet ^{o}C (^{o}F)	49 (121)	49 (121)	49 (121)
ΔT reheater ^{o}C (^{o}F)	36 (65)	33 (60)	33 (60)
Mist eliminator wash water rate, liters/sec (gpm)	0.8 (12.5)	0.9 (14.0)	0.8 (14.0)
Slurry flow to FDS, liters/sec (gpm)	137 (2170)	136 (2170)	88 (1400)
Slurry flow from FDS, liters/ sec (gpm)	83 (1317)	94 (1486)	NA

Table 3. YEARLY AVERAGE RELIABILITY FACTORS FOR CHOLLA FGD

Period	Reliability, percent		
	Module A	Module B	System Avg.
Jan. 74 - Dec. 74	94	88	91
Jan. 75 - Dec. 75	91	85	88
Jan. 76 - Dec. 76	89	89	89
Jan. 77 - Dec. 77	93	97	95
Jan. 78 - Nov. 78	98	98	98

Table 4. COST DATA FOR CHOLLA FGD SYSTEM

Period (year ended December 31 unless otherwise noted)	Total Operating and Maintenance Costs
1973	$ 74,600
1974	627,800
1975	339,000
1976	363,500
1977	359,000
1978 (Jan. 1 to Oct. 30)	313,998

NOTE: Includes: Operating labor and materials, maintenance labor and materials, limestone, and sludge disposal energy.

Excludes: Fuel differential charges and capital investment charges

Cost

The FGD system costs shown in Table 4 include operating labor and materials, maintenance labor and materials, limestone, and sludge disposal. However, these costs do not include fuel differential charges and capital investment charges.

Based on the original FGD system cost of $6.5 million, the capital investment charge amounts to about $1.5 million annually.

For a 115 MW unit this equates to an installed cost in 1973 of $56.5/KW. Currently, unit costs are estimated to be 1.5 to 2.0 times this cost with a resulting substantially higher annual capital charge. It should also be noted that capital charges for the new disposal facility, shared with Units 2, 3 and 4, are not included in the $6.5 million cited.

As has been pointed out by L. K. Mundth 1974 (1), the FGD system requires auxiliary supply of 2.8 MW of electricity and 18,000 pounds per hour of steam for reheat. These requirements are operational penalties reflecting in cost at the bus bar. Fuel differential charges are also properly assessed and average about $.8 million annually.

Maintenance Philosophy

During June, 1975, a preventative maintenance program was initiated on the Cholla I FGD system. As a part of this preventative maintenance program, the maintenance records of the FGD system components were analyzed to determine the frequency and nature of system component failures.

Based on this analysis, high maintenance components were identified and placed on a routine maintenance schedule. By this means, the number of emergency maintenance situations was reduced. Consequently, this also allowed a reduction in the overtime maintenance requirements.

This procedure has proven a successful approach as evidenced by the subsequent history of high reliability that the system has experienced.

SYSTEM COMPONENT ANALYSIS

Initial Major Problems and On Going Solutions

Following the initial startup of the Cholla I FGD system, several mechanical and chemical problems were encountered. The mechanical problems encountered during this shakedown period were as follows:

1. Vibration in the reheat sections due to improper flue gas distribution.
2. By-pass damper leakage due to distortions caused by flyash build up and prolonged exposure to high temperatures.
3. Malfunction of the flooded disc position control caused by binding due to build up around the disc shaft.
4. Booster fan leaks due to improper welding.
5. Erosion of the stainless steel pump impellers and liners.

In general these mechanical problems were resolved by means of equipment and operational modifications.

The excessive reheater vibration was eliminated by the installation of baffles in the duct work to improve the flue gas distribution.

The bypass damper leakage problem was alleviated by reducing the FGD system pressure drop so that a small amount of treated flue gas would flow backward through the dampers thereby reducing flyash buildup on the dampers.

However, no good solution was found for the erosion problems encountered by the stainless steel pump impellers and liners. These items must be replaced every six months.

In addition to the mechanical problems specified above, the following chemical problems were encountered:

1. Corrosion of the reheater.
2. Corrosion of expansion joints.
3. Failure of protective coatings.
4. Scaling of the first stage mist eliminator.
5. Scaling in the tangential nozzle area of the flooded disc scrubber.
6. Scaling in the flooded disc differential pressure sensing lines.

The reheater corrosion problem was the result of condensation in the ductwork leading to the reheater. This acid run-off caused tube necking at the reheater sheet resulting in tube failure. This problem was corrected by insulating the ductwork to prevent condensation.

The original metal expansion joints were replaced by a rubberized fabric type. However, it has been found that rubberized expansion joints wear out as a result of continued flexing. A complete solution to this problem has not yet been found.

Likewise, the protective coating failure problem has not been satisfactorily resolved. Although it was originally believed that the initial coating failures that occurred were due to improper coating application, subsequent reapplications of coating have continued to be unsuccessful on the B-side ductwork downstream from the reheater.

Scaling of the first stage mist eliminator has been controlled by redesigning the mist eliminator wash system for better coverage and frequency of washing.

However, solutions have not been found for scaling in the tangential nozzle area of the flooded disc scrubber or the flooded disc scrubber differential pressure sensing lines. The differential pressure sensing lines have been relocated to allow for more convenient access since these lines must be manually unplugged about every two weeks.

It has been observed that scaling problems tend to increase if the FGD system pH level is allowed to drop below pH 5.0. This is believed to result from increased oxidation of sulfites to hard-scale forming sulfates.

The pH control system has been improved by converting from the original in-flow type slurry sampling devices to a still-well type which eliminated sample line pluggage problems.

As the above discussions indicate, many improvements have been made on the original Cholla I FGD system.

EXTRAPOLATIONS FOR LIMESTONE THROWAWAY SYSTEMS DESIGN AND SELECTIONS

Process Loop Separation

One of the salient features of this system is the process loop separation. Arizona Public Service feels that the separation is dictated by the requirement for isolation of particulate removal from the principle SO_2 removal step.

This double-loop design affords some important advantages for its operation as a FGD system, both from the standpoint of process effectiveness and, also, from the standpoint of application of materials of construction.

As has been pointed out by Braden, 1978 (3), separation of the FDS (quench) loop from the absorber loop is important for isolation of the higher chlorides to the recycling slurry of the quench loop. This separation affords a more cost-beneficial selection of corrosion-inhibiting materials of construction between the two loops. Recent analysis indicate that the chlorides concentration of the quench loop 5 times the concentration of the absorber loop. In a later application of an FGD system at this station, chlorides in the quench loop are expected to rise to 10,000 to 14,000 ppm with application of cooling tower blowdown as make-up water and a tight station water balance. High molybdenum steels are indicated for use in this area with special attention paid to areas subject to abrasion.

Regarding reagent utilization, APS feels that the benefits of the double-loop design pays dividends. Operation of each loop at different and discrete pH levels provides for open-loop operation in the absorber tower and maximizes the full utilization of the reagent in the quench loop.

Open-loop operation of the absorber affords good scale control as well as mist eliminator cleanliness conditions. Isolated or closed-loop operation of the quench loop affords control of the rapid oxidation for enhancement of disposal products characteristics and is readily accommodated in the quench loop. Recent pilot tests on the Cholla 1 FGD system by Research-Cottrell gives an indication of potential for full oxidation by air sparging.

Materials of Construction

Liberal use of corrosion resistant metal alloys as compared to coated carbon steel construction characterizes the Cholla Unit 1 FGD system. The FDS scrubber, the absorber tower, and the reheater are 316L stainless steel construction based on results of the initial pilot test. Results of subsequent testing of materials in the system during operation is reported by Brodsky and Paul, 1975 (4), and indicate that extremes of service requirements are present. In the one extreme, only high-molybdenum alloys (over 6 percent) showed no local corrosion in the area of liquid-gas separation. In the least severe, most thoroughly washed mist eliminator area, only the sensitized 316 and 304 stainless steel evidenced wastage and pitting. Good service in the presence of acidic and chloride corrosion attack can be realized through use of nickle-based, high molybdenum alloys or with proper attention to the molybdenum content of the stainless steels and proper fabrication techniques.

We have had mixed success with the integrity of the remaining parts of the system which are constructed of carbon steel with corrosion resistant glass flake polyester-resin based linings. One disadvantage found for these materials is the lower tolerance to abrasion.

The specification of alloy metal vs rubber-lined process pumps has varied within APS for FGD systems depending upon the specific requirement of the application. All of the process pumps for the Cholla Unit 1 FGD system are alloy metal pumps, although APS experience with pump materials at our Four Corners Station demonstrated 500-10,000 hours of useful life for rubber-lined equipment, whereas all trial alloy metal pumps failed in 1000-1400 hours of service.

APS has found no panacea for selection of pump materials and certainly least of all first cost. Careful selection of available designs to satisfy the following major criteria is essential to good long-term satisfaction : (1) slurry abrasiveness, (2) combined corrosion/erosion mechanisms, (3) chloride/ pH factors, (4) head limitations, and (5) seal water requirements.

The state of the art design and experience with selection of materials for in-line reheaters in the pioneer era of the Cholla Unit 1 FGD system was basically experimental. The 316L stainless steel shell-and-tube reheaters of this unit have been satisfactory. Two factors have contributed to the success that has been achieved, i.e., split coil construction and adequate cleaning procedures. Contrariwise, reheaters of the same materials failed at our Four Corners Units 1, 2, & 3 application, because of difficulties arising from the in-place cleaning of unsectionalized coil reheater construcion.

In this regard, it should be noted that in the Cholla Unit No. 2 FGD system we have taken the plunge into Inconel 625 reheater materials for better long-term performance.

CLOSURE

In closing, there are two compelling observations regarding FGD system applications in general which is rooted in the Cholla experiences and which should be emphasized.

Cost-effectiveness for FGD systems which must be installed is heavily dependent on experienced design with emphasis on careful specification of process design and of materials selection for minimizing future operational and maintenance costs.

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Presented to NACE Annual Meeting (unpublished)

LA CYGNE STATION UNIT NO. 1

WET SCRUBBER OPERATING EXPERIENCE

by

Terry J. Eaton
Supervisor of Air Quality Control
La Cygne Station
Kansas City Power & Light Co.

Prepared For Presentation
at
EPA CONFERENCE
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MARCH 4 - 8, 1979

INTRODUCTION

This paper reviews the present operating experiences of the Babcock & Wilcox designed scrubber system and continues to describe the trends of costs, availabilities, modifications, manpower and other supportive data relating to operations since June, 1973.

Early operations proved the need for protective walls to house the modules; heavier rotor blades and improved bearings for the induced draft fans drafting both the boiler and scrubber; cyclone separators to prevent scale and debris from plugging slurry nozzles; developing fast scale removal techniques and systems to remove huge piles of debris; constant surveillance to repair internal corrosion damage; learning the economical trade-off for exotic materials to withstand abrasion, corrosion or violent operation; establishing the need for minimal instrumentation for consistent operation due to terrific maintenance requirements; and undoubtedly the most important accomplishment was to establish an operating force that proved a scrubber burning coal with a very high sulfur and ash could be made to work effectively. The major problems currently affecting the scrubber system are corrosion and lack of commercially available instrumentation to monitor critical parameters affecting the scrubber operation.

LA CYGNE STATION UNIT NO. 1

WET SCRUBBER OPERATING EXPERIENCE

DESCRIPTION

The 820-megawatt La Cygne No. 1 Unit began commercial operation on June 1, 1973, as a joint project of Kansas Gas and Electric Company and Kansas City Power and Light Company. The companies share equally in ownership and output and the unit is operated by KCP&L. The 630-megawatt No. 2 Unit, in service since being declared commercial May 15, 1977, operates under an identical arrangement.

The plant site is located about 55 miles south of downtown Kansas City, one-half mile west of the Missouri State line, and was selected based on locally available coal, water, and limestone. Construction of No. 1 Unit began in 1969 and erection of the Air Quality Control System was initiated in mid 1971.

Water for cooling purposes is furnished from a 2,600-acre reservoir constructed adjacent to the plant site. Fly ash and spent slurry from the AQC system is piped to a 160-acre settling pond located east of the reservoir.

Coal is delivered to the plant in off-the-road 120-ton trucks from surface mines operated by the Pittsburg & Midway Coal Mining Co. The nearby coal deposits are estimated to contain 70 million tons. The fuel is low grade, sub-bituminous with an as-fired heating value of 9,000 to 9,700 Btu/lb, and an ash content of 25 per cent and sulfur content of 5 to 6 per cent (Exhibit A).

Limestone is obtained from nearby quarries and delivered to the plant in off-the-road 50-ton trucks.

The boiler for No. 1 Unit is a cyclone-fired, supercritical, once-through, balanced-draft Babcock & Wilcox unit, with a rating of 6,200,000 pounds of steam per hour, 1,010 degrees F, 3,825 psig at the superheat outlet. The turbine-generator was supplied by Westinghouse and is rated at 874 MW gross output with five per cent overpressure and 3,500 psi throttle pressure. Three auxiliary, oil-fired boilers are used for plant start-up or for powering a 20 megawatt house turbine-generator. The net plant output is 820 megawatts, adjusted to include 24 megawatts used by the AQC system and 30 megawatts by plant auxiliaries.

PROCESS DESCRIPTION

The AQC system consists of eight two-stage Venturi-absorber scrubber modules (Exhibit B) designed to treat the boiler flue gas flow of 2,760,000 ACFM. (345,000 ACFM per module at 285 degrees F.) The ductwork design does not provide for flue gas bypass of the system. Also, the plant does not have an alternate or secondary fuel supply. Each module can be isolated for maintenance by individual dampers. On site limestone grinding and slurry storage facilities provide up to 1,000 tons of slurry per hour. The unit has a balanced draft system with three 7,000 hp forced draft fans and six 7,000 hp induced draft fans located between the AQC system and the 700 foot stack. There is a common plenum at both the scrubber inlet and outlet. Spent slurry and fly ash are removed from the module recirculation tank through rubber-lined pipes to the settling pond at the rate of 3,500 tons of solids per day. Clear make-up water is pumped from the pond and the loop is closed by recycling ball mill and module make-up water back into the system.

In abbreviated terms, as the hot flue gas enters the Venturi (Exhibit C), it is sprayed with slurry from 48 spray and 32 wall wash nozzles resulting in up to 99 per cent of the particulates agglomerated to the sump below. The gas continues through the sump making a 180 degree turn up through the absorber section. In the reaction chamber, the SO_2 is removed as the gas is forced through a limestone slurry solution sprayed on stainless steel sieve trays. The chemical reaction in part combines the calcium carbonate, water and sulfur dioxide to form two relatively insoluble calcium salts, calcium sulfate and calcium sulfite, which also fall to the sump. The cleaned gas passes through demisters to remove moisture and then is reheated to avoid deposits on the fans and provide buoyancy from the stack.

OPERATING EXPERIENCE

As a result of the continuing modifications and improved operating procedures, the module availabilities have steadily improved. The annual averages (Exhibit E) have been 31% for 1973; 76.3% for 1974; 84.3% for 1975; 92% for 1976; 92.5% for 1977; and 93.5% for 1978. With the addition of the eighth module in April 1977, continuous daytime load capability has exceeded 800 megawatts without appreciably affecting average module capability.

The results of a full load and stack emissions test on August 26, 1977, (Exhibit F) indicated module gas flow was still below crusing capability, the induced and forced draft fans were loaded well below rating and most systems were in good balance. Sulfur dioxide removal efficiency averaged 77% with individual modules averaging from 65 to 80%. Although particulate emissions from the plant have met EPA and Kansas State requirements, research and development work continues in an endeavor to reduce further the particulate emissions from Unit #1.

The ambient monitoring system continues to indicate ground level concentrations within the national standards for sulfur dioxide and nitrogen compounds (Exhibit H).

Limestone utilization has greatly improved with improved Ph control. In the past, it has been almost insurmountable to maintain inline glass cells without caking the limestone during shutdown or abrading the cells during operation with the high concentration of fly ash. By the proper maintenance discipline of acid flush, sonic cleaning and periodic water backflush, "straight line" Ph is resulting in approximately 30% less limestone, better control of scaling and has eliminated one more variable which hinders analysis in other areas.

Demister pluggage or scaling is no longer a problem at La Cygne. By eliminating the intermittent wash and moving the continuous wash (140 GPM) from below to above the first demister with increased number of nozzles (230 GPM), the chevrons operate "squeaking clean". Further experimentation may allow a reduction in these nozzles and perhaps sequential washing to reduce excess water.

Hard scale on the reheater tubes has been eliminated by the addition of a second layer of demisters in each of the modules. Scaling of the reheaters continues to be a problem, however it is soft and can be removed using fire hoses. The previous hard scale required high pressure water to remove the deposits.

MAINTENANCE

Cleaning schedules continue to call for taking one module out of service each night on a rotational schedule and keeping all modules available for the daytime peak loads. This allows a complete checkout of module internals to clean steam reheater pluggage, check nozzles for debris or loose rubber pluggage, to clean sump accumulation and to inspect for any other maintenance that could reduce reliability during the week. Module inspection and cleaning is not reduced to six hours or less with reheater pluggage the greatest problem. Water soot blowers may be the answer to cleaning on the line since steam blowers will help scale the carryover on the steam reheat tubes. Scaling is not one of our chief problems and we ordinarily ignore soft scale that forms on walls, on beams, or on the outside of nozzles.

Carryover to the induced draft fan blades continues to require regular washings. Each fan now requires cleaning once every four to seven days. Seldom is the high pressure wash necessary any longer, a "spinning" process with low pressure hoses has been very effective, cleaning the spare fan while out of service. The washings are usually done on a preventative basis, but must be taken out of service if bearing vibrations exceed 12 mils.

Rubber pipe linings and rubber-lined pumps have been an increasing maintenance problem. After several years operation, some materials that haven't been modified are wearing out. Rubber linings that tear out cause damage in other piping or pumps, plug nozzles and allow the steel pipes to wear through. Two or three years ago, this problem would not have been classified as serious, but this very abrasive slurry in practically continuous operation can be detrimental in trying to attain higher module availability.

Corrosion of carbon steel in the ductwork, dampers, induced draft fan rotors and housings, breeching and stack liner is and will continue to be our greatest concern. Burning extremely high sulfur coal and having the outage problems of a large unit creates periods of enormous SO_3 concentrations on these surfaces. This "cold end corrosion" damage requires extreme surveillance by maintenance engineers and unit outage plans must consider temperatures and time requirements for applying special coatings.

MANPOWER REQUIREMENTS

The scrubber operating and maintenance force is being increased to 54 people by adding one electrician for a total of two and two technicians for a total of three. The remaining personnel will remain the same (Exhibit I). The continued improvements in operating procedures and stable equipment operation should permit meaningful analysis of improved chemistry and control parameters. If the current effort to maintain Ph cells and SO_2 analyzers under challenging conditions are any indication, it will definitely require this increased force to make further progress.

Also worth noting are the increased demands on present maintenance personnel to accumulate, record and evaluate operating data on water saturation trends, limestone utilization, draft fan wear rates, reheater bundle failures, lined pump failures, rubber lined pipe replacements, nozzle replacements, spare parts, etc. The operators are also busy updating and extending operating instructions, special instructions and reviewing safety and training procedures.

COSTS

The total cost (Exhibit G) of the AQC system to date has increased to \$46.8 million or about 22 per cent of the \$213 million total Unit #1 cost, or about \$59 a kilowatt installed. It is estimated that an additional \$4 to \$6 million investment will be required to reach optimum system performance.

La Cygne Unit No. 1 production costs for 1977 for energy including coal costs average 6.54 mills per K.W.H. Production costs for the scrubber portion average 1.69 mills per KW. Discounting escalation, scrubber costs of labor and limestone are trending downward but maintenance materials have increased threefold.

Although the La Cygne Unit #2 has been commercially available since May 15, 1977, it is still too early to make cost comparisons between a scrubber system with local high sulphur - high ash coal and a precipitator only system burning Wyoming coal with greater transportation costs. Unit #2 has had a fantastic service record with 96% availability and 76 to 84% monthly load factors. It begins to appear that the local coal will be the most economical operation if probable western coal escalations are considered and installed costs of scrubber vs precipitator are not considered.

ADDITIONAL MODIFICATIONS

1. An improved steam source to increase the supply for module reheater service from 70,000 LB/HR to 120,000. This would then permit additional steam bundles for optimum reheat.

2. An additional sludge pond for deposit of scrubber spent slurry for approximately 30 years. A side benefit could be clear water recycled to the scrubber for improved chemistry.

3. Evaluate addition of third demister.

4. Make study on sub micron fly ash and sulphuric acid mist passing through scrubber without being collected. Although most scrubbers do not have required pressure drops for the duty, wetting agents, fogging arrangements or other developments could lead to a vast improvement.

5. Continue work to devise a better method to clean incline reheat tubes without taking equipment out of service.

LA CYGNE STATION
COAL AND ASH ANALYSIS

COAL

Proximate

Volatile	28.63
Fixed Carbon	37.94
Ash	24.36
Moisture	9.07
	100.00

BTU/lb. 9421

Grindability 59.59

Ultimate

Moisture	8.60
Carbon	51.93
Hydrogen	3.43
Nitrogen	0.94
Chlorine	0.027
Sulfur	5.39
Ash	24.36
Oxygen	5.33
	100.007

ASH

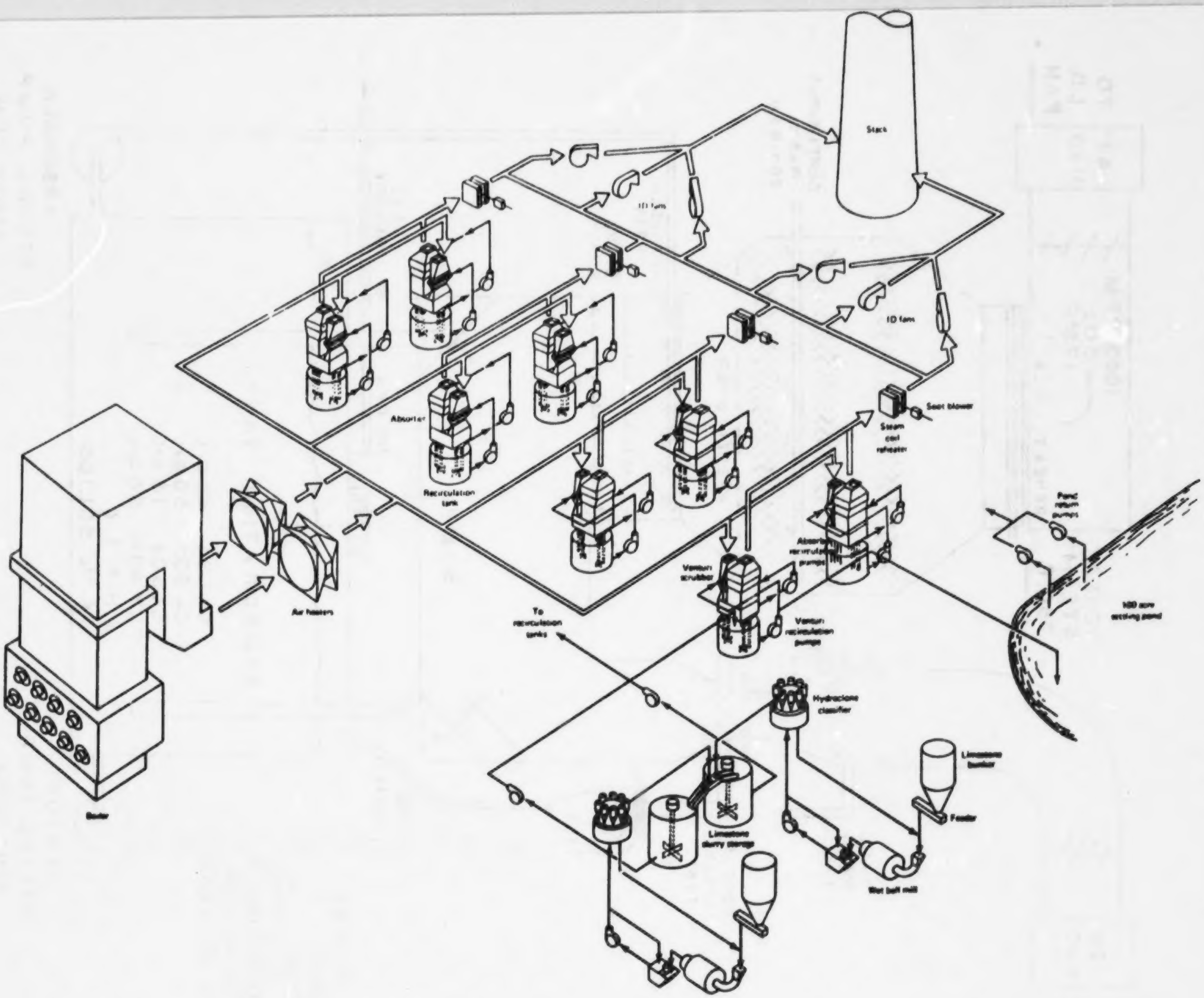
Analysis

Phosphorous Pentoxide	0.15
Silica	46.05
Ferric Oxide	19.23
Alumina	14.07
Lime	6.86
Magnesia	1.02
Sulfur Trioxide	7.85
Potassium Oxide	2.48
Sodium Oxide	0.60
Titania	1.02
Other	0.67
	100.00

Fusion Temperature

Reducing I.D.	1957
Soft (H=W)	2045
Soft (H=W/2)	2169
Fluid	2321
Oxidizing I.D.	2156
Soft (H=W)	2338
Soft (H=W/2)	2415
Fluid	2520

Exhibit A


Stack
ID fans
ID fans
Soot blower
Steam coil reheater
Absorber
Recirculation tank
Air heaters
Boiler
Absorber recirculation pumps
Venturi scrubber
Venturi recirculation pumps
Pond return pumps
160 acre settling pond
To recirculation tanks
Hydroclone classifier
Limestone slurry storage
Limestone bunker
Feeder
Wet ball mill

La Cygne limestone wet scrubbing system

Exhibit B

FIGURE 1 - LACYGNE FGD MODULE

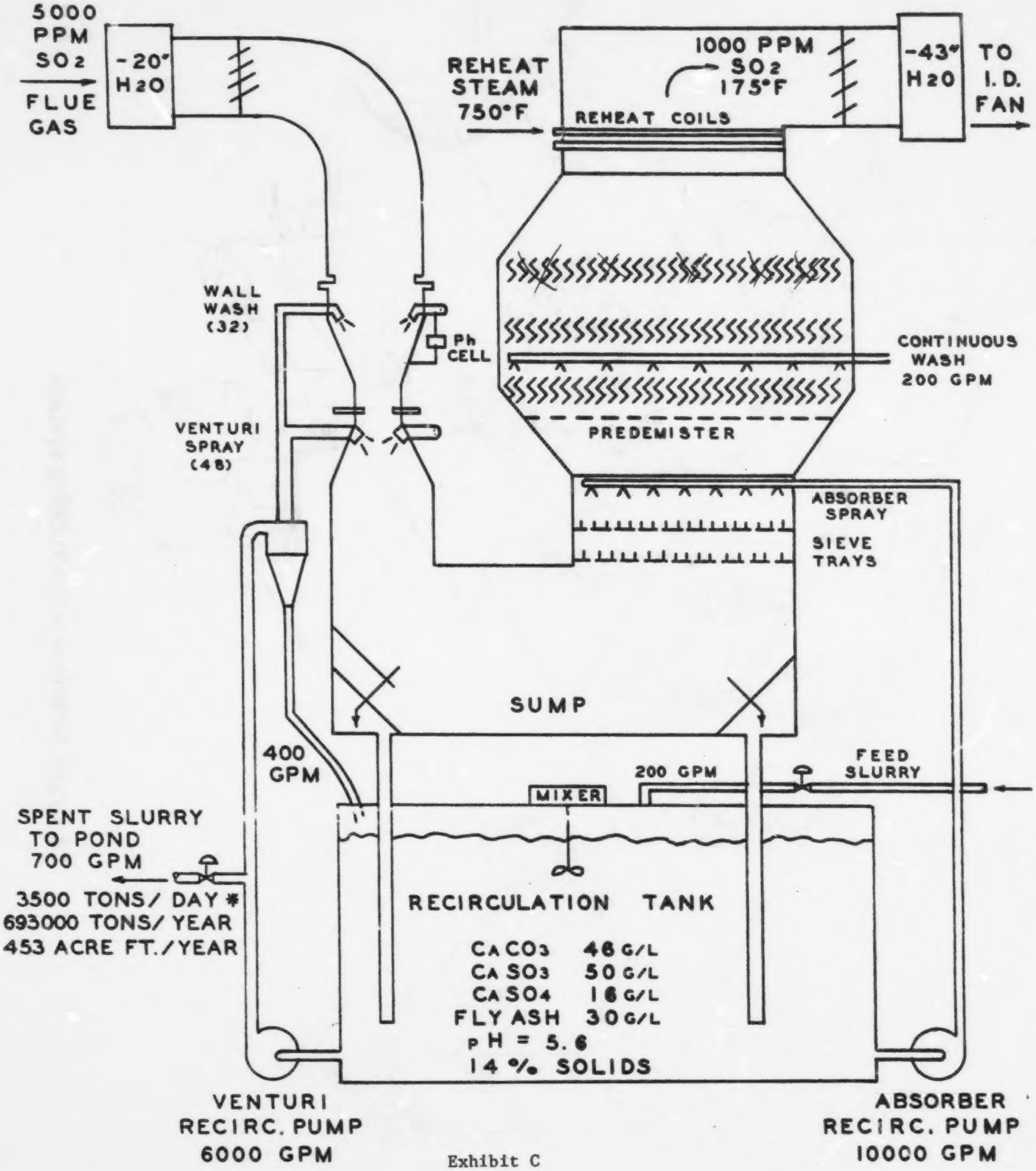

5000 PPM SO_2
FLUE GAS
-20" H_2O
REHEAT STEAM 750°F
REHEAT COILS
1000 PPM SO_2 175°F
-43" H_2O
TO I.D. FAN
WALL WASH (32)
Ph CELL
VENTURI SPRAY (48)
CONTINUOUS WASH 200 GPM
PREDEMISTER
ABSORBER SPRAY
SIEVE TRAYS
SUMP
400 GPM
200 GPM
FEED SLURRY
MIXER
SPENT SLURRY TO POND 700 GPM
3500 TONS/ DAY *
693000 TONS/ YEAR
453 ACRE FT./YEAR
RECIRCULATION TANK
$CaCO_3$ 46 G/L
$CaSO_3$ 50 G/L
$CaSO_4$ 16 G/L
FLY ASH 30 G/L
pH = 5.6
14 % SOLIDS
VENTURI RECIRC. PUMP 6000 GPM
ABSORBER RECIRC. PUMP 10000 GPM

Exhibit C

* TOTAL FOR ALL MODULES

1-10-78

LA CYGNE SCRUBBER WATER ANALYSIS

CATIONS	COOLING LAKE	SETTLING POND
CALCIUM (Ca)	126.4	808.0
MAGNESIUM (Mg)	16.3	106.0
SODIUM (Na)	31.0	52.5
POTASSIUM (K)	5.1	41.6
ANIONS		
BICARBONATE ALK (AS HCO_3)	112.2	79.3
CHLORIDE (Cl)	44.9	314.0
SULFATE (SO_4)	295.2	1995.1
SULFITE (SO_3)	* ND	* ND
SILICA (SiO_2)	1.12	52.0
OTHERS		
pH (pH UNITS)	7.7	7.5
CONDUCTIVITY IN MICROMHOS	820.0	3500.0
SOLIDS, SUSPENDED	5.0	5.0
DISSOLVED	610.0	3450.0

*ND - Not Detected

Exhibit D

MODULE AVAILABILITY SUMMARY - 1973

MONTH	A	B	C	D	E	F	G	AVERAGE % AVAILABILITY*	MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY											
FEBRUARY											
MARCH											
APRIL											
MAY											
JUNE	20	21	40	21	27	30	23	26	87,529	294	15.2
JULY	7	24	25	41	27	25	31	26	90,669	303	15.2
AUGUST	79	64	65	74	47	48	70	64	250,319	699	42.1
SEPTEMBER	13	0	13	13	13	0	0	7	20,073	95	3.5
OCTOBER	28	41	34	54	33	3	46	34	117,106	452	19.7
NOVEMBER	48	1	38	4	63	59	49	37	104,255	463	18.1
DECEMBER	42	20	5	31	26	11	32	24	61,013	339	10.3
								31%			17.7%

* MODULE HOURS
HOURS IN MONTH

Exhibit E

MODULE AVAILABILITY SUMMARY - 1974

MONTH	A	B	C	D	E	F	G	AVERAGE % AVAILABILITY*	MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY	49	32	44	87	23	37	81	50	35,862	364	6
FEBRUARY	66	68	59	76	52	100	65	69	85,256	364	16
MARCH											
APRIL	67	70	75	88	74	100	88	80	83,880	332	15
MAY	69	83	78	85	78	84	80	80	157,949	500	27
JUNE	92	84	83	90	82	83	87	86	185,473	480	32
JULY	75	80	80	81	85	79	77	80	110,122	313	19
AUGUST	90	90	73	81	81	78	99	85	231,382	571	39
SEPTEMBER	69	88	73	76	83	89	86	81	209,127	606	36
OCTOBER	71	61	59	81	79	93	89	76	230,302	662	39
NOVEMBER	90	71	60	61	84	85	84	76	130,128	386	23
DECEMBER											
								76.3%			25%

*MODULE HOURS
BOILER HOURS

Exhibit E (Cont'd)

MODULE AVAILABILITY SUMMARY

LA CYGNE 1975

MONTH	A	B	C	D	E	F	G	AVERAGE AVAILABILITY*	MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY		Turbine Generator Repair									
FEBRUARY		Turbine Generator Repair							7,886		
MARCH	82.4	96.03	89.5	76.6	92.96	91.5	96	89.33	244,873	694	41.1
APRIL		Generator Repair 25 Days							23,014		3.4
MAY	94.6	85.1	94.2	89.5	89.8	89.3	83.4	89.4	332,526	683	55.9
JUNE	87.8	85.4	83.9	84.9	84.1	86.1	88.6	85.8	324,952	667	56.4
JULY	78.4	89.7	89.6	83.7	85.4	87.4	85.2	85.6	297,870	590	50.0
AUGUST	74.64	88.07	87.29	78.01	92.44	85.00	83.06	84.07	294,402	630	49.5
SEPTEMBER	78.43	83.62	84.38	84.67	78.72	77.71	74.24	80.25	239,954	610	41.7
OCTOBER	66.16	77.26	46.27	73.62	71.91	73.07	64.69	67.57	74,660	231	12.5
		Generator Repair 15 Days									
NOVEMBER	92.87	90.79	80.18	93.18	96.09	89.39	93.94	90.83	165,058	346	28.7
		Generator Repair 17 Days									
DECEMBER	90.72	87.39	80.87	85.20	86.89	88.56	83.67	86.19	278,597	597	46.8
								84.3			38.6

*Working Hours + Reserve
Hours in Month

Exhibit E (Cont'd)

MODULE AVAILABILITY SUMMARY

LA CYGNE 1976

MONTH	A	B	C	D	E	F	G	AVERAGE AVAILABILITY*	MWH	BOILER HOURS	GENERATION LOAD FACTOR
JANUARY	85.8	84.6	90.7	71.8	83.9	82.3	84.3	83.3	301,641	620.5	50.6
FEBRUARY	93.9	90.3	85.8	91.2	91.7	93.1	94.6	91.5	308,361	594.5	55.4
MARCH	92.3	89.7	88.4	93.0	94.2	91.3	91.4	91.5	337,468	643.0	56.7
APRIL	92.3	90.5	88.7	97.1	95.8	98.0	94.8	93.9	76,810	143.0	13.3
		Scheduled Outage 24 Days									
MAY	96.5	92.1	93.5	95.7	89.4	95.3	96.2	94.1	223,048	436.3	37.5
		Scheduled Outage 9 Days									
JUNE	93.3	94.1	94.0	95.0	92.3	93.5	90.6	93.3	320,701	656.0	55.7
JULY	95.6	95.0	91.9	92.9	93.0	93.7	94.0	93.7	359,028	688.3	60.3
AUGUST	94.1	93.1	91.8	93.4	91.8	90.4	87.6	91.7	275,014	521.0	46.2
		Turbine Repair, Stack Relining 8 Days									
SEPTEMBER		Turbine Repair, Stack Relining 30 Days									
OCTOBER	97.4	96.7	97.5	89.0	96.1	96.1	96.1	95.6	88,925	255.8	14.9
		Turbine Repair, Stack Relining 18 Days									
NOVEMBER	94.7	93.3	93.7	95.3	94.2	91.3	93.6	94.0	342,236	626.8	59.4
DECEMBER	86.8	88.5	81.0	93.5	93.6	94.7	91.4	89.9	358,338	706.3	60.2
								92.0			46.4

$$*\frac{\text{Working Hours + Reserve}}{\text{Hours in Month}}$$

Exhibit E (Cont'd)

MODULE AVAILABILITY SUMMARY 1977

MONTH	A	B	C	D	E	F	G	H	Availability *	MWH	Boiler Hours	Generation Load Factor
JANUARY	94.2	90.0	95.0	95.1	94.5	91.6	89.8	---	92.9	255,822	539	43.0
FEBRUARY	93.4	93.0	92.6	93.8	93.3	93.9	88.0	---	92.5	310,748	590	57.8
MARCH	94.0	92.2	85.9	94.3	91.4	94.0	90.1	---	91.7	295,420	558	49.6
APRIL	96.1	93.7	97.0	94.2	95.2	96.1	94.5	---	95.2	178,226	384	30.9
MAY		GENERATOR REPAIR AND									15	
JUNE		STACK RELINING - 63 DAYS										
JULY	95.0	92.8	94.4	94.8	94.6	94.9	95.4	95.4	94.6	213,334	485	35.8
AUGUST	88.9	55.2	93.2	93.1	89.7	92.8	92.9	93.3	87.4	253,605	501	42.6
SEPTEMBER	93.2	93.7	89.1	90.0	92.8	95.0	91.7	93.0	92.3	287,701	524	49.9
OCTOBER	90.7	95.6	89.3	94.2	93.4	93.5	88.5	93.0	92.3	173,979	457	29.2
NOVEMBER	93.1	96.3	93.4	94.2	92.2	92.5	95.5	95.1	94.0	118,439	234	20.6
DECEMBER		TURBINE REPAIR Nov. 15 - Dec. 25										
									92.5%			39.9

Exhibit E (Cont'd)

*Working Hours and Reserve Hours
Hours in Month

MODULE AVAILABILITY SUMMARY 1978

MONTH	A	B	C	D	E	F	G	H	Availability*	MWH	Boiler Hours	Generation Load Factor
JANUARY	90.2	94.8	94.6	95.1	93.4	93.5	94.4	94.0	93.8	332,033	582	54.2
FEBRUARY	92.4	93.4	95.1	94.3	90.6	96.9	95.5	93.4	94.0	334,897	594	60.5
MARCH	95.3	95.2	90.4	95.4	94.4	94.7	88.6	93.3	93.4	264,961	593	43.2
APRIL	91.4	92.1	92.8	90.8	90.2	91.8	90.6	90.5	91.3	330,571	620	55.7
MAY	88.9	91.5	91.6	93.1	91.5	90.6	93.1	85.6	90.7	291,651	582	47.6
JUNE		O U	T A G E		6-8-78	thru 7-	17-78				14	0
JULY	87.5	97.2	91.9	93.9	88.4	92.8	93.1	95.3	92.6	160,847	340	26.2
AUGUST	92.1	92.5	95.0	95.7	92.7	94.3	94.7	95.3	94.0	307,378	579	50.1
SEPTEMBER	96.1	96.0	96.3	95.8	95.9	95.7	95.3	96.6	96.0	390,826	720	65.9
OCTOBER	95.9	95.5	96.3	97.0	97.0	97.6	96.7	96.3	96.8	138,126	255	22.5
NOVEMBER	91.7	94.9	94.3	93.3	93.6	93.0	94.3	96.1	93.9	386,402	720	65.1
DECEMBER	93.9	92.9	94.0	95.0	94.7	90.5	94.4	94.7	93.8	91,744	239	15
									93.5			42.2

Exhibit E (Cont'd)

*Working Hours and Reserve Hours
Hours in Month

LA CYGNE STATION UNIT NO. 1
FOUR HOUR FULL LOAD & STACK EMISSION TEST

DATE	August 26, 1977	NOX EMISSION:	0.81 # mm BTU
TIME	11:00 A.M. - 12:00 Midnight	AVERAGE SO_2 REMOVAL:	77%
LOAD RANGE:	800 + MW	PARTICULATE EMISSION:	.213 # mm BTU
AMBIENT TEMP:	94° F		

MODULES	A	B	C	D	E	F	G	H
GAS FLOW INDICATED	400	350	380	400	352	380	370	366
THROAT POSITION	OPEN	OPEN	OPEN	OPEN	OPEN	OPEN	OPEN	OPEN
REHEAT TEMPERATURE	170	190	150	190	185	180	160	170
VENTURI ΔP	5	5.5	5	5	5	5	5	5
REHEATER ΔP	2.5	5.5	4.5	4.5	5	2.55	4.5	5.5
ADSORBER DEM. ΔP	6.5	5.5	10	7.5	7.0	6.5	8.0	7.0
REHEAT OUTLET DAMPER POS.	50	100	96	38	100	52	100	100
ID FAN AMPS	380	420	380	400	470	470	(540 MAX)	
ID FAN INLET DAMPER POS.	42	42	32	36	36	40	(% OPEN)	
FD FAN AMPS	490	470	430				(540 MAX)	
LAB ph	5.45	5.7	5.55	5.7	5.58	5.77	5.72	5.29
SULFITE g/1	60.4	72.4	101.0	74.1	70.0	43.9	43.9	63.6
CARBONATE g/1	50.3	75.6	53.1	54.4	59.4	83.8	68.1	42.5
SO_2 EFFICIENCY %	80.0	82.1	74.9	64.3	76.4	72.1	73.1	74.8
INLET (PPM)	4600	4600	4600	4600	4600	4600	4600	4600
OUTLET (PPM)	920	825	1150	2285	1085	1285	1235	1160

CONDENSER VAC (IN. HG)	2.5	PRIMARY SUPER GAS PRESS. (IN.H_2O)	-8
WINDBOX FURNACE DIFF. PRESS (IN.H_2O)	32	HORZ REHEAT GAS PRESS. (IN.H_2O)	-9.5
SCRUBBER OUTLET PRESS (IN.H_2O)	-39"	ECON OUTLET GAS PRESS. (IN.H_2O)	-11.5
FURNACE PRESS (IN.H_2O)	-2	FEEDWATER PRESSURE (PSI)	4200
F.D. FAN DISCHARGE (IN. H_2O)	41	THROTTLE PRESSURE (PSI)	3400
PEND. REHEAT GAS PRESSURE (IN.H_2O)	-5	THROTTLE TEMP. (°F)	1000°
AIR FLOW (%)	85	HOT REHEAT TEMP. (°F)	1300
BOILER EXCESS O_2 (%)	2.2	FUEL FLOW %	68
BAROMETRIC PRESSURE (IN.Hg)	29.01	FUEL HEATING VALUE (MTB)	9800
STACK GAS TEMP (°F)	209	FLUE GAS VOLUME (MCFM)	2998
FLUE GAS MOISTURE (%)	13.66	STACK CO_2	13.4
STACK GAS VELOCITY Ft/Sec	103.15	STACK O_2 %	5.4

Exhibit F

COSTS

LA CYGNE STATION

Scrubber Operating Expense June-December 1973

OPERATING LABOR	$ 162,934	0.223 Mils/KWH
OPERATING MATERIALS	3,480	0.005
MAINTENANCE LABOR	189,400	0.259
MAINTENANCE MATERIALS	441,737	0.604
LIMESTONE	264,514	0.362
TOTAL	1,062,065	1.453 Mils/KWH

Scrubber Operating Expense 1974

OPERATING LABOR	284,541	0.223 Mils/KWH
OPERATING MATERIALS	67,032	0.053
MAINTENANCE LABOR	401,414	0.315
MAINTENANCE MATERIALS	335,486	0.263
LIMESTONE	780,297	0.613
TOTAL	1,868,770	1.467 Mils/KWH

Scrubber Operating Expense 1975

OPERATING LABOR	601,029	0.265 Mils/KWH
OPERATING MATERIALS	195,926	0.086
MAINTENANCE LABOR	416,206	0.184
MAINTENANCE MATERIALS	386,397	0.171
LIMESTONE	1,256,048	0.554
TOTAL	2,855,606	1.260 Mils/KWH

Exhibit G

September 30, 1977

LA CYGNE STATION

AMBIENT MONITORING SYSTEM

	STATION 1	STATION 2	STATION 3
MILES FROM PLANT	2	10	12.5
PRIOR TO START UP			
CONCENTRATION SO_2 - ppm	.009	.008	.003
NO_2 - ppm	----	.009	-----
RECENT LEVELS			
CONCENTRATION SO_2 - ppm		Overall	
1 hour high	.154	.653	.351
24 hour high	.098	.135	.106
Annual Average, Monthly	.015	.011	.016
	Station on Line and Wind Toward Monitors		
1 hour high	.119	.653	.209
24 hour high	.093	.135	.052
Annual Average	.014	.015	.016
	Station Shut Down or Wind Away From Monitors		
1 hour high	.154	.613	.351
24 hour high	.098	.041	.106
Average	.016	.010	.015
CONCENTRATION NO_2 - ppm		.023	
NATIONAL STANDARDS			
CONCENTRATION SO^2 - ppm			
Annual average		.030	
24 hour maximum		.140 (may exceed once/year)	
1 hour maximum		.500 (may exceed once/year- secondary standard)	
CONCENTRATION NO^2 - ppm			
Annual average		.050	

Exhibit H

LA CYGNE AIR QUALITY CONTROL
MANPOWER REQUIREMENTS

OPERATORS PER SHIFT

3 Attendants	13
3 Clean-Up	14
1 Shift Foreman	5
1 Process Attendant (Chemist)	1
	33

MAINTENANCE

Mechanics	8
Apprentice Mechanics	2
Welder	1
Electrician	2
Technician	3
Plant Helpers	2
Foreman	1
	19

ADMINISTRATIVE

Superintendent	1
Engineer	1
	2

TOTAL	54

Exhibit I

4H

DRY FGD SYSTEMS FOR THE ELECTRIC UTILITY INDUSTRY

Stephen J. Lutz, P.E.
TRW Environmental Engineering Division

C.J. Chatlynne, Ph.D.
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
Research Triangle Park

Much research is currently being directed toward the development and utilization of dry FGD technology because of its simplicity, lower energy requirements, as well as its ability to produce a dry, easy to handle product. There are several approaches to dry FGD, each with its own particular advantages and disadvantages. By comparing the current state of development for each of these technologies, we will attempt to provide the reader with the technical capability and comparative costs associated with each system, and give suggestions for additional research.

WHY CONSIDER DRY FGD?

In today's environment, we are experiencing an increasing public awareness of the various aspects of industrial pollution. Utility and industrial facilities are faced with a myriad of regulations governing atmospheric, water borne, and solid-waste discharges. If the U.S. is to remain competitive with foreign industries while continuing to be responsive to the environmental needs and concerns of our population, we must carefully evaluate the long-term impact of each of the various emission control techniques. FGD techniques vary widely in performance, reliability, and cost. Several of the newer approaches to flue gas SO_2 control technology additionally offer the capability of controlling several types of emissions in a unified fashion.

Dry FGD systems may offer several economic advantages over the current generation of wet FGD systems, but must be evaluated as part of an overall emission control approach. A dry FGD system providing control of both SO_2

and particulate emissions can be designed and constructed for a fraction of the cost of a comparable wet scrubber coupled to an electrostatic precipitator. Depending on the system, it may also provide a reduction in operating costs. A detailed comparison of these estimated costs will be provided later in this paper. In addition to the predicted cost savings, dry FGD systems will provide a reduction in energy consumption due to the elimination of the need for reheating the stack gas. The elimination of wet sludge, an emissions problem in its own right, will result from the utilization of dry FGD, but may be counter-balanced by a sodium salt leaching problem from the waste products of some of the dry processes.

WHAT IS DRY SORPTION?

Generally speaking, dry sorption refers to any process that directly produces a dry product. Usually one thinks of a baghouse using a dry SO_2 sorbent; however, also included in the category are processes that employ spray dryers followed by collection equipment such as baghouses, cyclones, or ESP's. For completeness, direct injection of sorbents into the boiler is included.

Baghouse FGD

Baghouse use is a simple approach in which the sorbent is either applied to the baghouse as a precoat or injected into the flue gas downstream of the air preheater. The latter technique is used to increase the residence time of the sorbent in the gas stream. Many sorbents have been used, among which are nahcolite and trona (naturally occuring $NaHCO_3$ and Na_2CO_3, respectively), $CaCO_3$, $Ca(OH)_2$, as well as commercial Na_2CO_3 and $NaHCO_3$. The sorbent reacts with SO_2 forming sulfite salts. The sorbent is periodically renewed so as to always have adequate reactive species in contact with the flue gas.

The following is a brief summary of non-EPA baghouse studies (additional details are presented later in this paper):

The first test using a baghouse for SO_2 control was at Southern California Edison's 320 MW Alamitos Station (1.5% S residual oil) in 1965. They reported successful use of dolomitic limestone for SO_2 removal; however, significant operating problems, which were not described in detail, were also mentioned. Nahcolite (natural $NaHCO_3$) was also used but not pursued due to lack of availability. The station has since been converted to natural gas (Bechtel, 1976).

Wheelabrator-Frye performed additional pilot-scale tests using a baghouse in 1967-1969 at Edwardsport Station of Public Service of Indiana. Many sorbents were examined, but only Na_2CO_3 and $NaHCO_3$ were found consistently effective. SO_2 removal ranged between 13 and 72%, and utilization between 22 and 93% (Bechtel, 1976). Unfortunately, high SO_2 removal was only possible at unacceptably low utilizations.

Air Preheater Company pilot tested several sorbents at Public Service Electric and Gas Company of New Jersey's Mercer Station in 1968-1969. They confirmed that lime is not an adequate dry sorbent, that nahcolite and commercial $NaHCO_3$ perform well, and that the operating temperature should be above 260°C (Bechtel, 1976).

In 1974, Superior Oil operated a bench-scale fixed bed of nahcolite at Public Gas Company's Cherokee Station. They observed 80 to 95% SO_2 removal at greater than 90% sorbent utilization (Bechtel, 1976).

Wheelabrator-Frye, Inc. tested nahcolite injection into baghouses at Colorado Ute Electrical Association's 11 MW Nucla Station in 1974. They determined that their best removal was 69% at a 56% utilization. The coal was mainly 0.8% sulfur with some testing on a 1.1% coal (Bechtel, 1976).

In 1976, the Electric Power Research Corporation (EPRI) contracted with Bechtel Corporation to survey the use of dry alkali for removing SO_2 from flue gas (Bechtel, 1976). Bechtel found that lime was relatively ineffective in a baghouse, that of 12 additional reagents, only $NaHCO_3$ and Na_2CO_3 appear to be sufficiently effective to warrant further consideration, and that higher flue gas temperatures generally result in better SO_2 removal. They concluded that baghouses injected with nahcolite appear to be the most promising method of removal in the dry state. Their study did not consider the use of spray dryers.

The following is a brief summary of EPA baghouse programs:

IERL-RTP's Particulate Technology Branch (PaTB) is co-funding a pilot system at Colorado Springs, Colorado, to examine the performance of nahcolite and trona on a 1000-1500 cfm (28-42 m^3/min) pilot baghouse (0.5% S coal). An option is available to expand this program to a full scale baghouse (80 MW).

A second PaTB project is on an industrial boiler (Kerr Industries) in Concord, N.C. Testing of sorbent regeneration is planned at Concord in addition to the basic sorption studies. Testing will be on two 35 acfm (1 m^3/min) baghouses (0.7 %S coal).

In 1977, TRW began a study of dry sorbents and fabric filters (Lutz et al., 1979). Their main conclusions were that dry sorbent baghouses exhibit economic advantages compared with current wet lime/limestone scrubbing processes when applied to western power plants burning low-sulfur coal. Additional conclusions are noted throughout this paper.

The main advantages of dry sorbent/baghouse systems are simplicity, energy requirement reduction, and the production of a dry, easy to handle product. The dry, once-through approach eliminates the complication of recycle and scaling that can occur in wet systems and, being dry, eliminates the need to reheat the flue gas. Possible disadvantages relate to the need to provide a hotter than usual flue gas (260°C) in order to achieve SO_2 removal in the range of 90% and, in the case of sodium-based sorbents, the need to dispose of soluble Na_2SO_3 in an environmentally acceptable manner.

Spray Dryers

An additional approach to increasing the contact time between the sorbent and the flue gas is to employ a spray dryer in which a slurry or concentrated sorbent solution contacts the flue gas and leaves the dryer as a dry powder. Collection is accomplished with an ESP, cyclones, or baghouses. Baghouses have the advantage of allowing additional contact between the flue gas and any unspent sorbent leaving the spray dryer. Sorbents used include $CaCO_3$, $Ca(OH)_2$, and Na_2CO_3.

There are currently three major commercial suppliers of spray dryer/baghouse systems: Rockwell International with Wheelabrator-Frye, Western Precipitator with NIRO (Lutz et al., 1979), and Babcock and Wilcox. Rockwell performed pilot tests at Basin Electric in 1977-1978. They (Estcourt et al., 1978) report 92% SO_2 removal at a stoichiometry of 1.0 using commercial soda ash (Na_2CO_3) in a spray dryer followed by a baghouse. They compare this with 74% for a baghouse alone with dry $NaHCO_3$ injection. Western Precipitator's spray dryer work was in Minnesota earlier this year. NIRO (Masters, 1978) reports SO_2 removal in excess of 90% at stoichiometric ratios as low as 1.3 to 1.5 with lime slurries in the spray dryer.

Two full scale systems have been sold in North Dakota and one in Wyoming: Western Precipitator at Basin Electric's 455 MW Antelope No. 2 (burning 0.68% S lignite), Rockwell at Otter Tail Power's 400 MW Coyote No. 1 (burning 0.9% S lignite), and Babcock & Wilcox at Basin Electric's 500 MW Laramie

River Station Unit 3. The first system is once-through soda ash while the latter two are lime/limestone systems; Rockwell also markets a calcium system. Current plans for the Rockwell system are for the spent sodium salts to be disposed of by burial in the ground, but details are unknown; however, regeneration a' la the Aqueous Carbonate Process is a future option. The first installations are scheduled to begin operation in mid-to-late 1981.

In addition to Rockwell International/Wheelabrator-Frye, Western Precipitator (Joy)/NIRO, and B&W, American Air Filter and Koch are also involved with spray dryers for FGD.

IERL-RTP's Process Technology Branch (PrTB) has initiated a full-scale demonstration of Rockwell International's Aqueous Carbonate Process at Niagara Mohawk's 10C MW Huntley Station in Tonawanda, New York, co-funded by the Empire State Electric Energy Research Corporation. This process utilizes sodium carbonate in a spray dryer to absorb SO_2 and has the added feature of regenerating spent Na_2CO_3 and producing elemental sulfur using coal directly as the reductant. A 2-year test program is expected to start in early 1982. Perhaps the main advantage of a spray dryer over a baghouse alone is the increase in sorbent residence time. When compared with lime/limestone scrubbing, any possibility of scaling is eliminated. It appears, also, that the use of a spray dryer is the only method of utilizing $CaCO_3/Ca(OH)_2$ in a dry process since these salts are not sufficiently reactive to be used in a baghouse alone. When utilizing a spray dryer for FGD, sufficient SO_2 removal is experienced upstream of the collection device to allow the use of collection devices other than fabric filters.

Direct Injection

The approach here is to inject the sorbent, limestone, directly into the boiler along with the coal. EPA investigated direct injection in the early 70's; however, boiler fouling caused the program to be discontinued. There do not appear to be any non-EPA programs; however, EPA is back in business, this time on a staged combustion, low NO_x burner. Apparently, the mechanics of combustion in the low NO_x burner eliminates the boiler fouling that had previously taken place.

The main advantage of direct injection is to minimize capital costs since a separate scrubber will not be required. One disadvantage is that high SO_2 removal efficiency has not been demonstrated; however, not being a capital-intensive process, it can be combined with other processes such as coal cleaning to achieve the required SO_2 removal.

OPERATING EXPERIENCE

Several projects have been undertaken, some of which are still in operation, and additional installations are planned for the near future. The greatest amount of experience has been obtained in baghouse-related testing, although the advancement of this technology is not progressing rapidly due to the current inability to obtain a sufficient supply of sorbent for full-scale demonstrations. Spray dry technology is currently the most advanced of the various dry FGD approaches, with several commercial units under development. Little operating experience exists with combustion-zone injection other than several EPA test burners. This technology is currently the furthest from commercial acceptability.

Combustion Zone Injection

No commercial applications of combustion-zone injection for dry FGD are planned. Table I lists the relevant operating experience with this technology. Of particular note are the positive results demonstrated recently using dry limestone as a sorbent material. These tests, conducted by the IERL-RTP's Combustion Research Branch, involve the mixing of ground limestone with the coal under combustion conditions designed to minimize NO_x formation by controlling the temperature/stoichiometry history of the reactants. This combustion condition provides prolonged reactant residence times under fuel-rich conditions and lower peak flame temperatures. Currently available preliminary data from this program are encouraging but not conclusive and will be subject to verification by additional testing.

TABLE I

COMBUSTION ZONE INJECTION
OPERATING EXPERIENCE

FACILITY	DATE	SORBENTS TESTED	COMMENTS
EPA Test Burner	Early 1970's	Dry limestone	The test program was not successful due to recurring problems with injection system.
Superior Oil Test at Cherokee Station at Public Service of Colorado	1974	Nahcolite, commercial sodium bicarbonate, predecomposed sodium bicarbonate, soda ash	Test was run on a pilot boiler burning No. 2 fuel oil. Nahcolite was the most efficient for SO_2 removal with commercial sodium bicarbonate, predecomposed sodium bicarbonate, and soda ash being less efficient. A major finding of these tests was that nahcolite and commercial sodium bicarbonate particles exploded (thermal comminution) because the CO_2 and water formed by decomposition could not be liberated fast enough.
EPA Low NO_x Test Burners	Current	Dry limestone	IERL-RTP's Combustion Research Branch is currently running tests on three small coal-fired test burners (30 kW, 30 MW, 300 MW heat input). Although the primary purpose of these tests is to evaluate a distributed mixing burner (DMB) for minimizing NO_x formation, limestone addition to the coal was also evaluated for SO_2 control. The limestone was found to have an apparent effectiveness, providing greater than 50% reduction in SO_2 with a Ca/S mole ratio of 1. No clogging or slagging was experienced.
EPA Pilot Plant (Fluidized Bed Combustion)	Current	Limestone	IERL-RTP's Advanced Process Branch is currently evaluating limestone injection for SO_2 control in FBC technology. The pilot plant is a 38 cm x 38 cm bed with a gas flow of 600 scfm (17 m^3/min). Test results are not yet available.

Current testing is limited to several EPA test programs. Three small coal-fired experimental burners (30 kW, 30 MW, 300 MW heat input) are available. Additional testing is being performed in conjunction with the fluidized bed combustion (FBC) program. Early operating experience with this dry FGD process resulted in numerous operating problems, including materials handling difficulties, boiler slagging, and generally poor performance. The current EPA testing program represents an advance in the design of the sorbent addition system and, based on the results from the initial phase of testing, appears to have overcome these operating difficulties. It must be understood, however, that these tests are pilot plant scale studies and may not be representative of operating characteristics found in full size utility boilers.

Baghouse FGD

The development of dry FGD baghouses has fostered research in two distinct areas: the evaluation of possible sorbent materials, and ~~the~~ development of techniques to utilize dry sorbents in practical operating systems. The earliest testing, conducted by Southern California Edison at the 320 MW Alamitos Station (1965), Wheelabrator-Frye at Edwardsport (1967-1969) (pilot), and Air Preheater at Mercer Station (1968-1969) (pilot), was primarily concerned with the evaluation of sorbents. The development of dry FGD baghouse technology has proceeded through several test programs: Wheelabrator-Frye at Nucla Station (1974) and Basin Electric (1977), and by KVB (1978); and through several major engineering studies (Bechtel, 1976; Lutz et al., 1979). Table II reviews the relevant operating experience with this technology.

Operating experience with these test systems has been satisfactory but, because they have all been designed for discrete testing, no long-term continuous reliability data have been available. Baghouses have, however, been used in the electric utility industry for control of particle emissions over a considerable time period and have been accepted as reliable. The addition of a dry sorbent material to the collected ash is not expected to significantly decrease the reliability of these devices.

TABLE II

BAGHOUSE FGD
OPERATING EXPERIENCE

ACILITY	DATE	SORBENTS TESTED	COMMENTS
Southern Cal. Edison	1975	Dolomitic limestone nahcolite	Significant operating problems were experienced at this facility. Testing was discontinued due to these problems and the inability to obtain an adequate supply of nahcolite.
Wheelabrator-Frye at Edwardsport	1967-1969	Sodium bicarbonate, soda ash, potassium permanganate, calcium hydroxide, and 12 others	This was a side stream pilot scale test which ran for 2 years. SO_2 removals ranged to 72% with utilizations of 22-93%. Due to the small scale of this pilot program, the operating experience is not considered applicable to full-scale installations.
Air Preheater at Mercer Station	1968-1969	Commercial sodium bicarbonate, nahcolite, and hydrated limes	Tests were run on an existing baghouse which had been used for particle removal testing. Baghouse flow varied from 7,500 to 15,000 cfm (200-400 m^3/min). Considerable operating experience was obtained using a variety of injection techniques.
Wheelabrator-Frye at Nucla	1976-1977	Nahcolite	16 independent 90 minute tests were run on a small (11 MW) coal fired unit. Flow rate was 65,000 scfm (1800 m^3/min).
Wheelabrator-Frye at Basin Electric	1976-	Nahcolite	Various system configurations were tested over a 4-month period.
KVB Bench Test	1978	Commercial sodium bicarbonate, trona, nahcolite	Bench scale test demonstrating the Buell-Horbid baghouse design.

EPA's current and planned testing of the dry FGD baghouses is limited to two test programs.

IERL-RTP's Particulate Technology Branch has initiated a baghouse program at an 80 MW generating system owned by the City of Colorado Springs, Colorado. The fuel used is 0.5% sulfur western coal. The chief program objectives of baghouse evaluation and dry sorbent injection will be accomplished with a 1000-1500 cfm (28-42 m^3/min) pilot baghouse, fully instrumented, with a typical air/cloth ratio of 2/1 (cfm/ft^2 or 0.3 m^3/min per m^2) of filter. The current testing program is planned for 15 months. The contract contains an option for continued testing of the pilot baghouse, and/or the construction of a full scale baghouse (80 MW) which would run an additional 15 months. The utility itself has recently decided to install a pilot-scale spray dryer.

The second test program currently underway is the expansion of an IERL-RTP industrial baghouse project to evaluate SO_2 sorbents. The location is at Kerr Industries in Concord, N.C. The installation includes a 35,000 acfm (100 m^3/min) baghouse on each of two 60 MW boilers, using a 0.7 to 0.8% sulfur coal. The air/cloth ratios (m^3/min per m^2) of filter for the baghouses are capable of ranging from 1/1 to 3/1; 2/1 is a typical operating ratio. The dry sorbent injection studies are scheduled to begin by spring 1979, using a number of sorbents and a range of operating conditions. Further work is being contemplated that would include a regenerable sorbent process. The industrial boiler project may be the most likely candidate for regenerable sorbent investigations, since higher sulfur fuel usage will generate more solid waste when once-through sorbent methods are used.

No commercial applications of dry FGD baghouses are contemplated in the near future. Nahcolite, the most reactive of the dry sorbent materials, is not currently available in quantitites necessary for a full-scale baghouse installation, and the owners of nahcolite reserves are reluctant to open a commercial-size mine (at least 500,000 tons per year).

Spray Dryers

Spray dryers, a combination of a spray contactor and a particle collector, rely heavily on the operating experience obtained from traditional approaches to particle collection (baghouses, cyclones and electrostatic precipitators), and from some of the newer FGD research programs, including the dry sorbent baghouse pilot demonstrations and the spray contactor wet scrubbing systems. The combination of these individually proven technologies has been demonstrated in a relatively few pilot plant test programs, identified in Table III, which have yielded valuable results. Basing their designs on these test results, Rockwell International, Western Precipitator, and B & W have each sold one full scale installation. Testing on these installations should provide adequate performance, reliability, and operating data from which other utilities can evaluate their options for flue gas desulfurization. In addition, IERL-RTP and the Empire State Electric Energy Research Corporation are co-funding the demonstration of Rockwell International's Aqueous Carbonate Process--a regenerable Na_2CO_3-spray dryer process that produces sulfur.

DRY SCRUBBING AND PROPOSED NEW SOURCE PERFORMANCE STANDARDS

The NSPS revisions proposed in the September 19, 1978 Federal Register for steam-electric generating facilities include a requirement of 85% SO_2 removal (24 hour average) with maximum emissions of 520 ng/J (1.2 lb/million Btu). In addition, the percentage removal will not apply if SO_2 emissions are reduced to 86 ng/J (0.2 lb/million Btu).

Recently, because of what appears to be some distinct advantages in meeting the revised SO_2 standards, a great deal of interest has developed in dry scrubbing processes. It appears that these processes may offer improved reliability and reduced capital and operating costs for selected applications. Among these applications are sources burning fuels with sulfur contents which would allow meeting the 0.2 lb/million Btu limit with less than 85% removal treating all or a portion of the flue gas. Dry sorption has the potential for "full" scrubbing where less than 100% of the flue gas is

TABLE III

SPRAY DRYER FGD
OPERATING EXPERIENCE

FACILITY	DATE	SORBENTS TESTED	COMMENTS
Rockwell International and Wheelabrator-Frye at Leland Olds Station	1977-1978	Soda ash, lime, flyash, flyash/lime mixtures	Pilot plant testing for 2 months to obtain scale-up data for future installations.
Western Precipitator at Hoot Lake Station	1978	Soda ash, trona, pot-ash, limestone, lime, flyash	Pilot plant testing on 20,000 acfm (570 m^3/min) at 150°C to obtain design data for later applications.
Babcock & Wilcox	Continuing	Lime, trona	Pilot test program utilizing Hitachi Ltd. and B&W spray dryers. A new, 120,000 acfm (3400 m^3/min) prototype is under construction.

scrubbed. Additionally, a dry sorbent fabric filter system also removes particulate matter, and some dry sorption approaches conserve energy because the flue gas temperature is not appreciably lowered, nor is the gas saturated as in wet scrubbing. These advantages result in potentially simpler, less energy consumptive (due to minimization or elimination or reheat) and less costly systems for compliance with NSPS.

COSTS

Despite all analyses of the technical and environmental advantages of one system over another, the choice of FGD systems by an electric utility usually boils down to answers to just two questions: (1) Which system will satisfy the EPA? and (2) What does it cost? The capital and annualized operating costs for each of the dry FGD systems is compared with a similarly designed wet scrubber (limestone) in Table IV.

Combustion Zone Injection

Of the various dry FGD techniques discussed in this paper, combustion zone injection offers the greatest potential for cost savings. Current EPA research is progressing into a design which will control NO_x, SO_2, and particulate matter with a single, low capital investment design. Although much research is needed before a definitive cost analysis can be performed, the primary components of this system appear to be a system for mixing the dry sorbent material with the coal upstream of the pulverizers, a specially designed burner, and a particle collection device. Capital costs are expected to compare very favorably with baghouse FGD or spray dryer systems.

Operating costs should reflect the use of inexpensive sorbents, such as limestone, and the only utilities required will be for sorbent transport and particle collection. Operating labor should be minimal since there will be

TABLE IV

COMPARATIVE COSTS[a] FOR SELECTED FGD SYSTEMS 500 MW BOILER

		COMBUSTION ZONE INJECTION	BAGHOUSE FGD	SPRAY DRYER	WET LIMESTONE
Capital Costs:	$\$10^6$	Unknown	23	43	59
	$/kW		46	86	118
Annualized Operating Costs:	$\$10^6$/yr	Unknown	9.1	10.9	14.3
	mills/kWh		2.6	3.1	4.1

[a] In 1977 dollars.

no new process systems such as injectors or contactors to require maintenance and, because of the totally dry nature of this system, a maximum amount of flue gas sensible heat can be recovered by air preheaters.

Baghouse FGD

Although no full-scale systems have been sold at this time, sufficient data have been obtained from the extensive dry sorbent testing programs to establish accurate cost parameters for baghouse FGD systems. TRW's Environmental Engineering Division has developed a FGD baghouse design for a 500 MW pulverized coal-fired utility boiler (Lutz et al., 1979) based on typical utility requirements.

Spray Dryers

Three full-scale spray dryer FGD systems have been sold to date. A comparison of costs for the spray dryer versus a wet scrubber has been made for the Laramie River Station Unit No. 3 and will be presented in a later paper at this symposium (Janssen, 1979).

Wet Scrubbers

Comparison costs were established for wet scrubbing by the TVA, utilizing a computer cost analysis program modeled on their Shawnee wet limestone scrubber (Torstrick et al., 1977).

FUTURE RESEARCH NEEDS

Further development of the various dry FGD techniques is anticipated and can be expected to increase performance and reliability and decrease costs of future designs. Some of this research will be funded by the EPA, primarily in the areas of developing technologies. Competition among the various suppliers of commercial dry FGD systems is expected to reduce costs through design improvements as more of these systems are sold.

Combustion zone injection FGD technology is still in its infancy. The EPA is providing some basic research in this field and will continue to develop it in future programs. One interesting possibility is the application of combustion zone injection FGD to the fluidized bed combustion concept now being developed.

Dry sorbent baghouse FGD technology has developed to the point where the basic FGD process is well understood. Commercialization has not been forthcoming due to the unobtainability of nahcolite in sufficient quantity. Research needs to be established into possible ways to reduce or eliminate the large consumption of this material by increasing utilization. Sorbent regeneration appears to be the most likely candidate but must be proven in pilot-scale testing before it can be applied to any commercial installations. Other research needs associated with the dry sorbent baghouse FGD system include further development of methods of insoluabilization of the waste sorbent so that it can be disposed of in an environmentally acceptable manner.

Spray dryer FGD technology has advanced at a very rapid rate and is now available commercially. Improvements in reducing the costs for these systems can be expected but will require extensive development. Regeneration of sorbents offers great promise in this area, but will require significant research and development before it is practical on a commercial scale.

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PLAN, DESIGN AND OPERATING EXPERIENCE OF FGD FOR COAL FIRED BOILERS OWNED BY EPDC

by

Yasuyuki Nakabayashi

Assistant General Manager

Thermal Power Department

Electric Power Development Co., Ltd.

Tokyo, Japan

ABSTRACT

EPDC has built Wet Limestone-Gypsum FGD systems for coal fired boilers which are comprised of five 250MW class capacity, totaling 1,280MW.

Each of the FGD systems has been in service for three to four years with the same operating reliability as the boiler availability. EPDC is planning and constructing two 500MW and one 700MW coal fired plants where EPDC is planning to use imported coal and install FGD Systems accordingly.

EPDC is planning to start project of another 1,000MW coal fired plant in the near future.

This paper is presented to describe basic philosophy for design, specific features on design, improvement of design resulted from operating experience, operating reliability, major problems and treatment of by-product.

PLAN, DESIGN AND OPERATING EXPERIENCE OF FGD
FOR
COAL FIRED BOILERS OWNED BY EPDC

1. BASIC PHILOSOPHY OF FGD SYSTEM

It is indispensable to evaluate in advance which type of process is the most feasible for application of FGD.

The following factors are to be evaluated in advance

- o Technical reliability
- o High performance capability
- o Good economy
- o No secondary pollution
- o Superior operating characteristics
- o Security of stable supply of absorbent and disposal of by-product

The optimum FGD process shall be selected on basis of the total evaluation on above six factors.

As EPDC judged Limestone-Gypsum process as the best one, EPDC adopted the process.

2. EPDC'S PRE-EVALUATION

The following is an outline of EPDC's philosophy as to how the Limestone-Gypsum process has been evaluated.

(1) Selection of Wet Process or Dry Process

At the time when EPDC was planning to adopt FGD system (1967 - 1968), dry processes were developed in Japan with Government Subsidy.

EPDC concluded that wet processes were superior to dry processes on the basis of pre-evaluation of the six factors previously mentioned after a complete study of the development status of dry processes and the details of new technical development on wet processes.

Table 1 show its outline.

Table 1. Comparison of Wet and Dry FGD Processes

Item	Wet FGD Process	Dry FGD Process
Desulfurization efficiency	90 percent or more (Influence of variation of flue gas volume on efficiency is very small.)	Around 80 percent (Efficiency varies according to volume of flue gas or operating hours)
Temperature of treated exhaust gas	50° to 60°C (Reheat is required for prevention of white plume and better diffusion to the atmosphere)	Reheat is not necessary as boiler flue gas temperature is sufficiently high.
Material of equipment	Mainly plastic lined material (Anti-corrosive measures are needed)	Mainly metal material (Heat-proof and anti-corrosive measures are needed)
Scaling up	Easy	Difficult
Application to coal-fired boiler	Suitable	Not suitable because of high dust concentration
Reaction velocity	Fast	Slow
Investment cost	Small	Large
Equipment	Small (smaller space is required)	Large (larger space is required
Pressure loss	Small	Large
Consumption of utilities		
Water	Much	Little
Absorbent	Cheap (limestone)	Expensive (activated carbon)
Electricity	Much	Little
Fuel	Much	Not required
Steam	Much	Not required

(2) Selection of the Optimum Process out of Wet Processes

The wet process can be sub-classified to several ones depending on the kind of absorbent and the treatment method of by-product.

Each process possessed advantages and the disadvantages, respectively, however, the Limestone-Gypsum process has been evaluated as the best one. Table 2 summarizes the pre-evaluation results.

Table 2. Pre-evaluation of Wet FGD Process

Evaluation Item / Name of Process		Process			Facility of Operation	Actual use	Cost	
		Desulfurization Efficiency	Simplicity	Secondary Pollution			Construction	Operation
Throw-away Process	(1) Calcium base	○	◎	△	◎	○	◎	○
	(2) Sodium base	◎	◎	△	◎	○	◎	△
Gypsum recovery Process	(3) Lime-Limestone	○	○	○	○	◎	○	○
	(4) Double Alkali	◎	△	○	△	○	△	○
	(5) Sulfuric acid Dilution	○	○	○	○	○	△	△
	(6) Ammonia-Calcium	◎	△	△	△	△	○	△
Sulfuric acid recovery Process	(7) Sodium base	◎	×	○	×	○	×	○
	(8) Magnesium base	◎	×	◎	×	○	×	○

Remarks ;
◎ Better
○ Good
△ Bad
× Worse

3. OUTLINE OF EPDC'S FGD SYSTEM

Since EPDC started operation on February 2nd, 1975, of EPDC's first FGD system at Takasago Power Station No.1 unit, EPDC has adopted FGD Systems to all of coal fired power plants owned by EPDC. All EPDC's FGD Systems can treat the full capacity of boiler flue gas and the total capacity is equivalent to 1,280MW generating capacity in terms of facilities. The FGD process is that which uses limestone as absorbent and produce gypsum as by-product.

The boiler fuel is mainly the coal which is produced in Japan.

The FGD systems are the extension of facilities to those of existing power stations, Table 3 shows the outline of power stations and those FGD systems.

Table 3. Outline of Design Specifications

Item			Isogo Thermal P/S		Takasago Thermal P/S		Takehara Thermal P/S
		Power Station / Unit	No.1 Unit	No.2 Unit	No.1 Unit	No.2 Unit	No.1 Unit
Location			Yokohama City, Kanagawa Pref.		Takasago City, Hyogo Pref.		Takehara City, Hiroshima Pref.
Generation Equipment		Approved Output (MW)	265	265	250	250	250
	Turbine	Type	Cross compound 4-Turbine Chamber, 4-Shunt, Suction, Reheating System.		Reheating, Regenerating, Circulation System.		Tandem, 3-Turbine Chamber, 4-Shunt, Suction, Reheating system.
		Output (MW)	265	265	250	250	250
		Manufacturer	Tokyo Shibaura Electric		Mitsubishi Heavy Industries		Hitachi
	Boiler	Type	IHI-FW Single Drum Radiation Reheating Water Pipe System.		Forced Circulation Reheating Readiation Water Pipe System.		Hitachi B & W Natural Circulation Single Drum Radiation Reheating System.
		Max. Evaporation (t/h)	840	840	825	825	810
		Manufacturer	Ishikawajima-Harima Heavy Industries		Mitsubishi Heavy Industries		Babcock-Hitachi Hitachi
	Fuel	Fuel	Coal	Coal	Coal	Coal	Coal
		S Content in Fuel	0.6	0.6	1.8	1.8	2.0
		Fuel Consumption (t/h)	100	100	96	96	100
		Operation Started on	May 25, 1967	Sep. 23, 1969	Jul. 1, 1968	Jan. 18, 1969	Jul. 25, 1967
Flue Gas Desulfurization Equipment		Type	Wet Limestone-Gypsum Process		Wet Limestone-Gypsum Process		Wet Limestone-Gypsum Process
		Volume (Nm^3/h)	900,000	900,000	842,000	842,000	852,000
		Volume (MW Equivalent)	265 (Whole Energy)	265 (Whole Energy)	250 (Whole Energy)	250 (Whole Energy)	250 (Whole Energy)
		Absorbent	Calcium Carbonate		Calcium Carbonate		Calcium Carbonate
		Manufacturer	Ishikawajima-Harima Heavy Industries		Mitsui Miike Works		Hitachi
	Waste Water Treatment	Type	Coagulating Sedimentation plus adsorption		Coagulating Sedimentation plus adsorption		Coagulating Sedimentation plus adsorption
		Volume (t/h)	15	15	15	15	15
		Operation Started on	Mar. 3, 1976	May 21, 1976	Feb. 5, 1975	Mar. 24, 1976	Feb. 1977

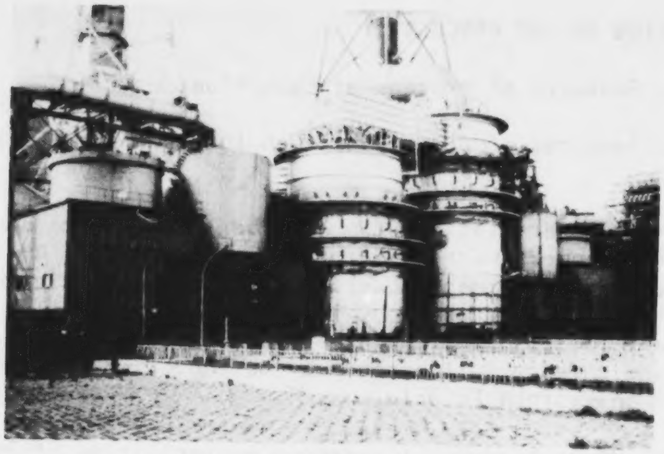


Photo 1 Isogo's FGD

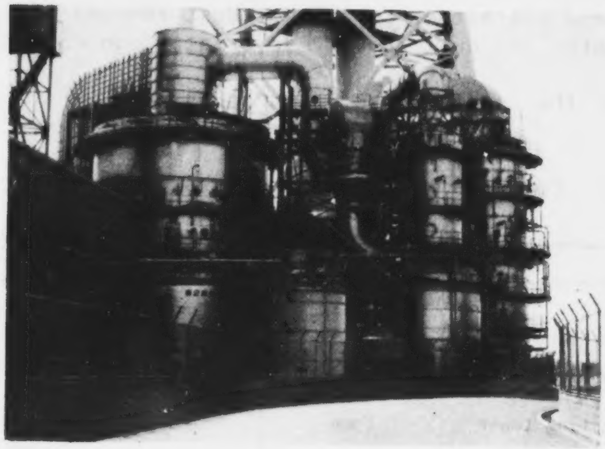


Photo 2 Takasago's FGD

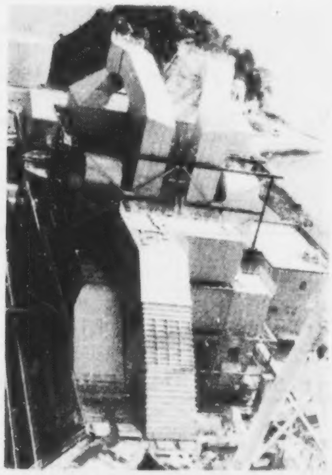


Photo 3 Takehara's FGD

4. DESCRIPTION ON FGD PROCESS

(1) Specific Features of Process at Each Plant

Specific features of process at each location are described as follows:

At Takehara Power Station, two scrubber towers were adopted in parallel and cooling towers were used for removal of dust. At Takasago Power Station, a PH adjustment tower is installed for adjustment of PH of slurry from first scrubber by introducing part of flue gas into the PH adjustment tower so that high desulphurization efficiency may be maintained and unreacted limestone may be converted to gypsum without applying any sulphuric acid.

At Isogo Power Station, the receiving facilities of limestone and gypsum are small because of low SOx concentration in flue gas. The FGD system is designed and guaranteed for higher dust removal efficiency because emmission control of dust is very sever at Isogo Power Station.

Table 4 shows the comparison of FGD Processes.

Table 4. Comparison of FGD Processes in Actual Operation

	Isogo	Takasago	Takehara
Scrubber	2 series	2 series	2 parallel
Cooling tower	None	None	2
PH Adjusting tower	None	1	None
Absorbent excess ratio	1 ∿ 1.1	1 ∿ 1.1	1 ∿ 1.1
H_2SO_4	little	None	little
SOx in flue gas	300 ppm	1500 ∿	1500 ∿
Eliminator washing water	Raw water	Raw water & mother water	Raw water
After burner (gas temp.)	used (85 ∿ 95°C)	used (80 ∿ 85°C)	used (120°C)

(2) Description of the Process at Each Plant

(a) Process description of IHI's FGD system at Isogo P/S

Flue gas boosted up by fan is sent to the first scrubber, then, the gas goes through the second scrubber. SOx contained in boiler flue gas is removed in the first and second scrubbers and the dust in it is also removed in the same scrubbers.

The wet gas flowing through the second scrubber is mixed in the mixing chamber with the hot air warmed in the reheater, and is discharged from the stack after heated up by the after burner to a temperature of 85°C ∿ 95°C. The absorbent is 98% pure limestone with a particle size of minus 325 mesh.

The absorbent is carried by container trucks for exclusive use at the power station and is stored in a silo.
After weighing, it is kept in a tank. After measuring the concentration, the volume of absorbent slurry is automatically determined to match with the same volume of flow from the scrubber.

The amount of feed is about 1.0 ∿ 1.1 times of the inlet SO_2 by molecular weight.

The circulation slurry from second scrubber is supplied to the first scrubber.

The bleed slurry from the first scrubber is fed to the oxidation tower and then the slurry is oxidized with normal pressure air for production of gypsum.

The 5% gypsum slurry after oxidation is fed to a gypsum separation apparatus.

After concentrating the slurry by a thickener, the slurry is processed by centrifuges for production of 10% moisture cake.

The product is stored in a gypsum warehouse and a part of the thickner overflow water is discharged out of the FGD system in order to prevent build up of such impurities as Cl, F etc. The mist eliminator is washed intermittently with industrial raw water. Figure 1 shows the flow chart of Isogo's FGD.

(b) Process description of Mitsui Miike's FGD System at Takasago P/S

Flue gas boosted up by about 450 mmAq by use of fan is split into three divisions for introduction into 1st and 2nd oxidation towers (No.2 Unit has only one oxidation tower), pH adjustment tank and 1st scrubber. Then, the gas goes through 2nd scrubber for completion of SO_2 removal, dust removal and pH adjustment. The wet gas coming through the 2nd scrubber is mixed in the mixing chamber with the hot air generated in the reheater, and is discharged from the stack after heated up to a temperature of 80°C ∿ 85°C.

The absorbent of limestone of purity over 98% and particle size of minus 325 mesh is received from a ship for exclusive use and

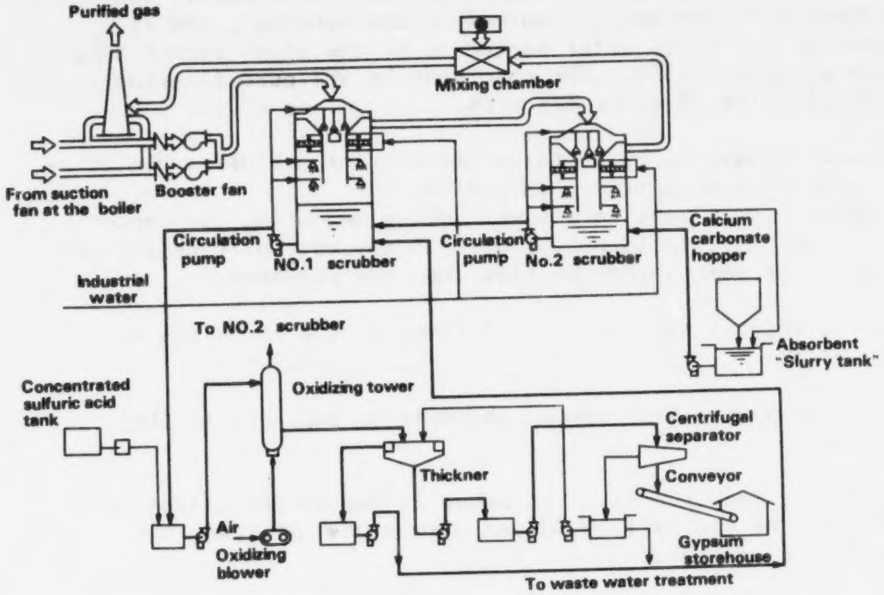

Purified gas
Mixing chamber
From suction fan at the boiler
Booster fan
Circulation pump
NO.1 scrubber
Circulation pump
No.2 scrubber
Calcium carbonate hopper
Industrial water
To NO.2 scrubber
Absorbent "Slurry tank"
Concentrated sulfuric acid tank
Oxidizing tower
Centrifugal separator
Thickner
Conveyor
Air
Oxidizing blower
Gypsum storehouse
To waste water treatment

Figure 1 The Flow Chart of Isogo's FGD

some container trucks, and stored in silo. After weighing, it is kept in the tank as about 15% slurry. After measurement of density, volume of the absorbent slurry is automatically determined to meet the volume of flue gas treated, inlet concentration of SO_2 and pH of oxidation tower. Then, it is fed into 2nd scrubber. The amount of the feed is about 1.0 ∿ 1.05 times of inlet SO_2 by molecular weight.

The 2nd scrubber is being operated at pH 6.0 ∿ 6.3 and liquid/gas ratio 6.0 ∿ 7.0 ℓ/m^3. The circulation slurry from the 2nd scrubber is supplied to the 1st scrubber. The 1st absorption tower is being operated at pH 5.6 ∿ 1.1 and liquid/gas ratio 5.5 ∿ 6.0 ℓ/m^3.

The pH value of the bleed slurry from the 1st scrubber is adjusted to pH 5.4 ∿ 5.8 in the pH adjusting tower. In the oxidation tower, the air is blown into the bleed slurry and the slurry is oxidized under normal pressure for gypsum production.

A very small quantity of catalyst solution is supplied to the 2nd scrubber to improve the SO_2 removal, to increase the oxidation speed and to obtain quality gypsum.

The 5 ∿ 7% gypsum slurry after the oxidation is sent to gypsum separating apparatus. After concentration to about 20% slurry by the thickner, this slurry is processed by the 7 centrifugal separators to produce cakes of 5 ∿ 8% moisture. The product is stored in the gypsum warehouse.

The thickner overflow containing 500 ppm or less suspended solids and the filtrate of centrifugal separators are circulated for adjustment of absorbent density, adjustment of liquid level of the scrubber and for washing of mist eliminator. A portion of the liquid is discharged out of the system for the purpose of preventing accumulation of Cl^-.

The mist eliminator of a 4 stage chevron type is being washed intermittently by use of raw water and circulating mother liquor.

Figure 2 shows the flow chart of Takasago's FGD.

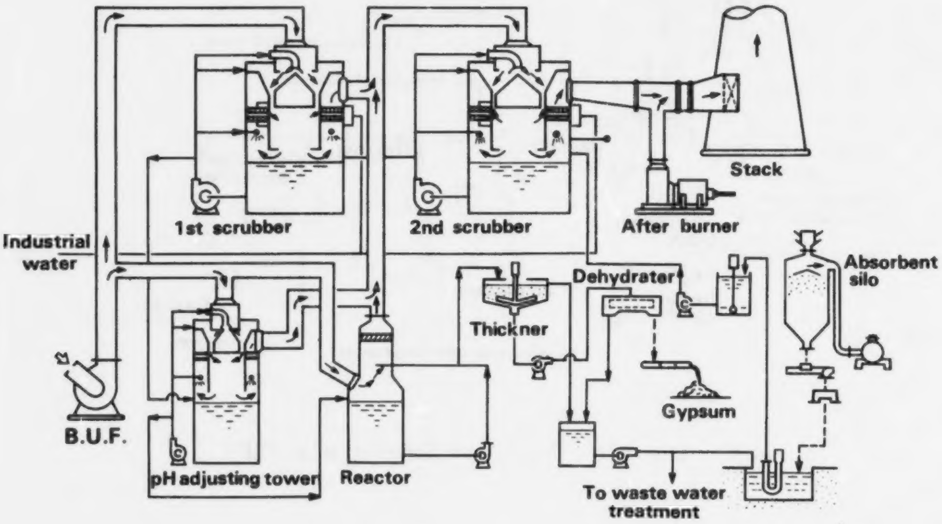

Stack
1st scrubber
2nd scrubber
After burner
Industrial water
Dehydrater
Absorbent silo
Thickner
B.U.F.
Gypsum
pH adjusting tower
Reactor
To waste water treatment

Figure 2 The Flow Chart of Takasago's FGD

(c) Process description of Hitachi's FGD system at Takehara P/S

Flue gas boosted up by use of fan is split into two divisions, A cooling tower and B cooling tower.

In the case of Takehara P/S, the FGD is composed by 2 trains.

Flue gas temperature is reduced at the cooling tower from 150°C to 50°C ∿ 60°C and the flue gas is sent to scrubber. Dust is removed at the cooling tower and SOx is removed at the scrubber.

Limestone as the absorbent is the same specification of the others, and unloading system of limestone and Gypsum Production System is the same as Isogo's process.

Figure 3 shows the flow chart of Takehara's FGD.

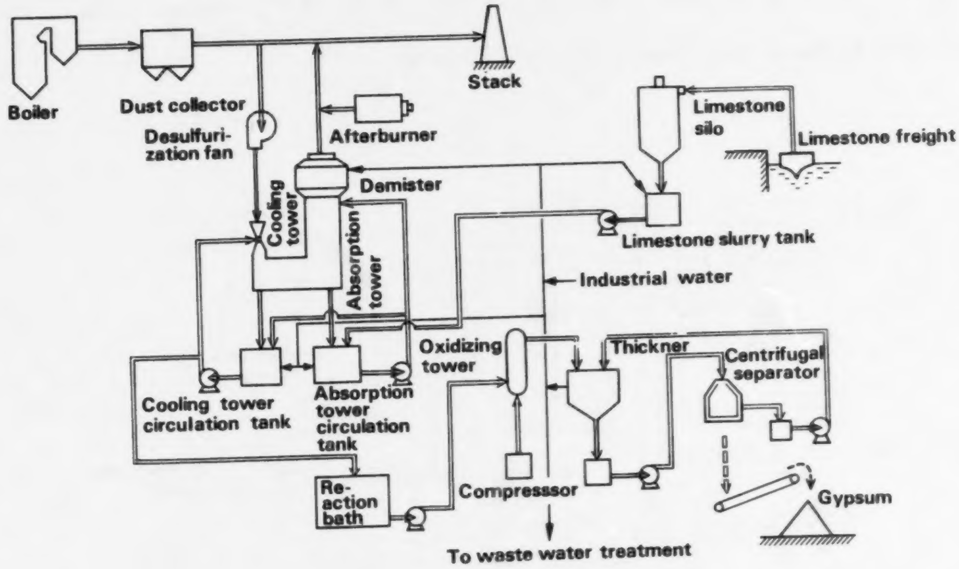

Boiler
Dust collector
Stack
Desulfuri-zation fan
Afterburner
Limestone silo
Limestone freight
Demister
Cooling tower
Absorption tower
Limestone slurry tank
Industrial water
Oxidizing tower
Thickner
Centrifugal separator
Cooling tower circulation tank
Absorption tower circulation tank
Re-action bath
Compresssor
Gypsum
To waste water treatment

Figure 3. The Flow Chart of Takehara's FGD

5. SPECIFIC FEATURES OF DESIGN AND OPERATION OF FGD SYSTEM

There are a number of factors to be considered in the design and operation of FGD system in order to keep a high desulphurization efficiency and maintain the equipment in good conditions.

(1) Specific Features of Design

The following specific features were considered in the design of EPDC's FGD systems.

(a) Specific features of process

- The limestone absorbent is purchased in the form of powder with a size of minus 325 mesh directly from the limestone production mine. There are no crushing facilities of wet mill for FGD process at the power plant site for.

  The reasons following:

  - o Problem exists in the reliability of wet mill crushing facilities
  - o FGD systems must be installed within the limited area of power stations as the FGD system is the extension of existing facilities
  - o There are crushing facilities already at limestone mine so it will be double investment
  - o Satisfactory countermeasures are needed for prevention of noise generated from the crushing facilities
  - o Quality control (purity and particle size) was quite possible by purchasing in a form of powder and there is no influence to the desulphurization efficiency

- Production facilities of gypsum as by product are installed in power stations.

  Following is the reason:

  - o Gypsum is a marketable item and sales is possible
  - o Gypsum is a stabilized chemical material.

- The scrubber is of the Venturi type which possesses high dust removal capability

  Following is the reason:

  - o Limitation is found in the capability of electrostatic precipitators
  - o Severer regulations apply to dust emission

(b) Consideration for corrosion

- Epoxy lining or rubber lining is applied in the scrubber and its connecting pipes and pumps

Following is the reason:

- o PH in the scrubber line is less than 7.
  Corrosive slurry moves in the line so direct contact to metallic surface must be avoided.
- o Epoxy material or Titanium metal is used for the parts where lining is not possible.

(c) Coordination with the plant

- A flue gas by-pass duct was installed for

  the following reasons:

  - o No influence to the boiler is caused by a B.U.F. trip
  - o No necessity is found for flue gas to flow through the FGD system at the time of plant start up.

(d) Waste water treatment

- Waste water discharged from FGD system is treated.

  Following is the reason:

  - o FGD waste water must be treated as PH, SS and COD etc. exceeds the limit of Japanese waste water standards
- Specific COD treatment system is provided for FGD system

  Following is the reason:

  - o With a conventional COD system, the COD materials discharged from FGD system cannot be treated satisfactorily and therefore the specific COD treatment system was developed for FGD system.
- Calcium substances are used as a coagulate sedimentation agent for treatment of fluorine

  Following is the reason:

  - o Treatment of heavy metal in FGD waste water is made by such coagulate sedimentation agents as Na-substances and Ca-substances. Fluorine cannot be treated by Na-substances.

(2) Consideration on Operation

Following consideration shall be paid for operation of FGD system

(a) Continuous blowing water out of the system

Following is the reason:

- o Coal contains chroline and some tens ppm chroline exists in flue gas. Chroline is feared to concentrate in the scrubber which causes corrosion of the equipment, therefore the constant chroline concentration must be kept with the continuous blowing water.

(b) Good control shall be achieved on the concentration of absorbent slurry

- o Desulphurization efficiency is influenced by particle size of absorbent and PH of absorbent slurry.

  The particle size is almost constant as the absorbent is purchased in the form of powder.
  The PH control of absorbent slurry is achieved satisfactorily by a good control of slurry concentration as it changes considerably depending on the input amount of absorbent.

- o The desulphurization efficiency goes up for the increase of input amount of absorbent and it causes the scale build up, therefore the optimum input amount exists.

(c) The amount of absorbent is determined according to plant load

- o Unless the absorbent is supplied following the load change, the material balance of absorbent which contributes to reaction with SOx will be broken then the desulphurization efficiency will be affected.

(d) Sulphur content in the coal shall be kept constant

Following is the reason:

- o If the sulphur content in the coal changes considerably, the pH control according to the change of sulphur content shall become impossible as the mass in FGD system is too large even though the matching amount of absorbent is fed accordingly.

Therefore the variation of sulphur content shall be controlled to the minimum and a certain range of sulphur content coal must be used when different coals are blended.

(e) Raw water shall be applied for washing eliminator

Following is the reason:

- In case of desulphurization of high sulphur content flue gas, it is often feared that scale build up takes place at the Venturi and mist eliminator of the scrubber and further it will cause plugging.

  If the process water from dehydrator is utilized for the washing water of eliminator, the scale build up occurs but it was prevented by changing the process water to raw water.

6. OPERATING RELIABILITY

The operating reliability of the FGD system can be judged by the following factors.

- Availability of FGD system has equalled the availability of plant
- Maintained high performance
- Stable operation

Assurance has been obtained on the statisfactory reliability of FGD systems which EPDC is operating, based on the past operation experience of FGD systems.

Operating experience is described as follows.

(1) Operating Hours

Table 5 shows the operating hours of each unit at each plant since the start of operation through fiscal year end of 1977.

Following facts have been proven.

- Total operating time of all FGD systems came to approximately 80,000 hours which are equivalent to 99.3% of availability of EPDC's plants.
- The longest operating time per unit is approximately 25,000 hours and its availability is 98.7%. (Takasago No.1 unit)
- 100% is the availability of FGD systems at Isogo No.1 and No.2 and Takehara No.1 for the fiscal year of 1977.

The above data were taken from the figures through March 31, 1978 and FGD systems have been in service at present without any problem.

The estimated operating time will be as follows by March 31, 1979.

- o Total availability time: Approx. 120,000 hours
- o The longest availability time per unit: Approx. 30,000 hours

Table 5. Actual Hours of Plant and FGD Operations

[Unit: hour]

Plant	Unit	1974		1975		1976		1977		Total		Avail-ability	Remarks
		A	B	A	B	A	B	A	B	A	B		
Isogo	1			676	676	7,699	7,699	8,115	8,115	16,490	16,490	100 %	o 1974 means period of April 1, 1974 to March 31, 1975. o A means hours of Plant operation. o B means hours of FGD operation.
	2					7,031	7,031	7,641	7,641	14,672	14,672	100 %	
	Total			676	676	14,730	14,730	15,756	15,756	31,162	31,162	100 %	
Takasago	1	975	957	8,206	8,053	7,515	7,438	8,381	8,328	25,077	24,776	98.7%	
	2					8,143	8,008	7,703	7,595	15,846	15,603	98.4%	
	Total	975	957	8,206	8,053	15,523	15,446	16,084	15,923	40,923	40,399	98.6%	
Takehara	1					1,056	1,056	7,895	7,895	8,951	8,951	100 %	
Total	-	975	957	8,924	8,729	31,741	31,252	39,735	39,574	81,036	80,512	99.3%	

(2) Average Desulphurization Efficiency

Table 6 indicates average desulphurization efficiency by the fiscal year end of 1977 according to plant and unit.

Following facts have been learned.

- o The average desulphurization efficiency is about 91% through approximately 80,000 hours of FGD operation.

- The average desulphurization efficiency at Isogo Power Station is about 96% through approx. 30,000 hours of FGD operation.

- The average desulphurization efficiency at Takasago Power Station is about 88% through approx. 40,000 hours of FGD operation.

- The average desulphurization efficiency at Takehara Power Station is about 91% through approx. 9,000 hours operation.

Table 6. Actual Average Desulphurization Efficiency

[Unit: %]

Plant	Unit	1974	1975	1976	1977	Average	Remarks
Isogo	1		94	96	96	96	Year means fiscal year
	2			96	96	96	
	Total		94	96	96	96	
Takasago	1	90	87	88	90	88	
	2			88	90	88	
	Total	90	87	88	90	88	
Takehara	1			91	91	91	
Average	-	90	88	91	92	91	

(3) The Load Following Capability of FGD System to the Load Change of Power Plant

According to a load pattern, EPDC's coal fired power plants are in service at full load during day and at about 1/2 load during night, but at the minimum load in the week end due to reduced demand.

This load pattern for a week is repeated. The following capability of FGD system to load change is dependent upon how to control the pH value in the scrubber at constant level.

To maintain a constant desulphurization efficiency during load swing, the feed rate of absorbent into the scrubber is controlled in proportion to the produced gypsum quantity thus the PH value is kept constant.

As a result of investigation of desulphurization efficiency for random consecutive 60 days in 1978, according to load change of the plants at three locations, the variation of desulphurization efficiency is summarized in the following Table 7.

Table 7. The Variation of Desulphurization Efficiency

[Unit: %]

Plant	Load Variation	Max. $DeSO_x$ Efficiency Variation	Min. $DeSO_x$ Efficiency Variation	Average $DeSO_x$ Efficiency Variation
Isogo	265MW ∿ 135MW	1.2 (95.8 ∿ 97.0)	0 (97.4 ∿ 97.4)	0.1 (96.8 ∿ 96.7)
Takasago	250MW ∿ 125MW	3.9 (90.8 ∿ 94.7)	1.7 (93.3 ∿ 95.0)	2.7 (92.2 ∿ 94.9)
Takehara	250MW ∿ 100MW	8.4 (88.0 ∿ 96.4)	0.8 (93.4 ∿ 94.2)	2.9 (93.1 ∿ 96.0)

The following facts have been learned.

- o Difference exists in the variation according to FGD system at each plant.
- o As the sulphur concentration in flue gas changes considerably and desulphurization efficiency is affected accordingly, the efficiency is not always assured to be more than 90%.
- o The average desulphurization efficiency value has increased compared to those of past years. The increase has been realized due to the efforts exerted in developing EPDC's know-how of the absorbent control line.

(4) The variation of Desulphurization Efficiency at Load Change

Figure 4 shows the variation of desulphurization efficiency at the time the plant load decreases from 250MW, rated value, to 125MW at Takasago Power Station.

The load changes by 3 MW/min. so it takes about 42 minutes to reduce 250MW down to 125MW.

Figure 5 shows the change of SOx and O_2 on Chart Recorder, O_2% increase along with load down from 250MW to 125MW so SOx concentration at the inlet of FGD system decrease.

SOx concentration at the outlet of FGD system decreases accordingly, the desulphurization efficiency changes in an almost constant range without showing unstable characteristics.

The stable desulphurization efficiency is maintained after load change has been completed.

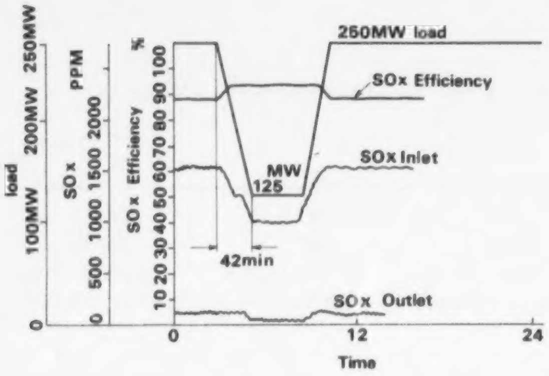


Figure 4. The Variation of Desulphurization Efficiency at Load Change

As for SOx concentration, no calibration has been made based on O_2 concentration.

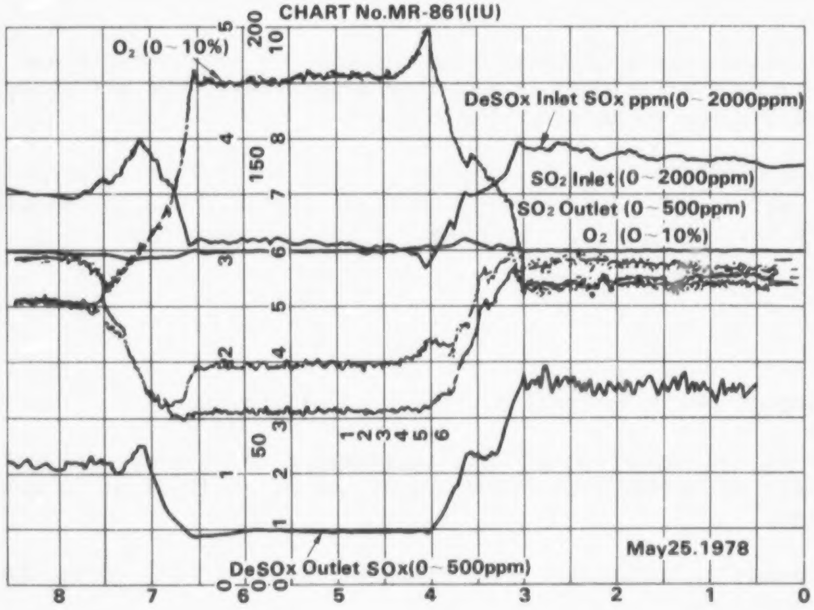


Figure 5. The Change of SOx and O_2 on Chart Recorder

7. CHANGE OF SULPHUR CONTENT DUE TO BLENDING OF DIFFERENT COALS

(1) Method of Blending Coals

Three to four kinds of coals are blended for boiler fuel at each power station, and the variation range of sulphur content is intended to be as small as possible.

As for a blending method, the coals are stockpiled in the predetermined area of power station, according to the kind of coal then the certain coals are scrapped by bulldozers and thrown into a underground hopper where several kinds are blended.

The variation range of sulphur content becomes as follows.

Location	Actual data	Blend ratio	Sulphur content
Isogo	0.5 ∿ 0.6%	A:B:C = 8:1:1	A 0.3% B 2.5 ∿ 2.7% C 0.6 ∿ 1.0%
Takasago	1.6 ∿ 1.85%	D:B:A = 4.25:4.25:1.5	D 1.0 ∿ 1.4% B 2.5 ∿ 2.7% A 0.6 ∿ 1.0%
Takehara	1.5 ∿ 1.9%	D:B:A = 3:6:1	D 0.9 ∿ 1.5% B 2.1 ∿ 2.7% A 0.3 ∿ 0.5

8. MAJOR IMPROVEMENT ON DESIGN

A problem arose in the basic design of FGD system at Takasago Power Station during its operation as scale was built up in Venturi and eliminator of the scrubber.

The scale build up was monitored by an increase of pressure drop, and was manually removed, stopping the FGD system if necessary.

This has been the major trouble of FGD system and there have no other troubles in the operating experience.

Following is the description on the solution of scale build up at Takasago Power Station.

Overflow water (mother liquid) from a thickener was utilized for the washing water of eliminator at the time the plant operation started. Since $CaSO_4$ is saturated in the mother liquid, it tends to be crystalized as scale onto the eliminator in the scrubber.

Following improvement was performed in June 1978.

- Raw water was supplied in addition to the mother liquid to the washing water line of eliminators of the first scrubber and the second scrubber. (Takasago P/S)
- It is desirable to change the mother liquid to raw water also for the washing water of eliminators, but it was found to cause further burden

to the existing waste water treatment system because the volume of blowing water out of the system must be increased due to the balanced volume of total water.

With the modification, the scale build up on the eliminators was considerably prevented.

Figure 6 shows the improvement of eliminator washing water line.

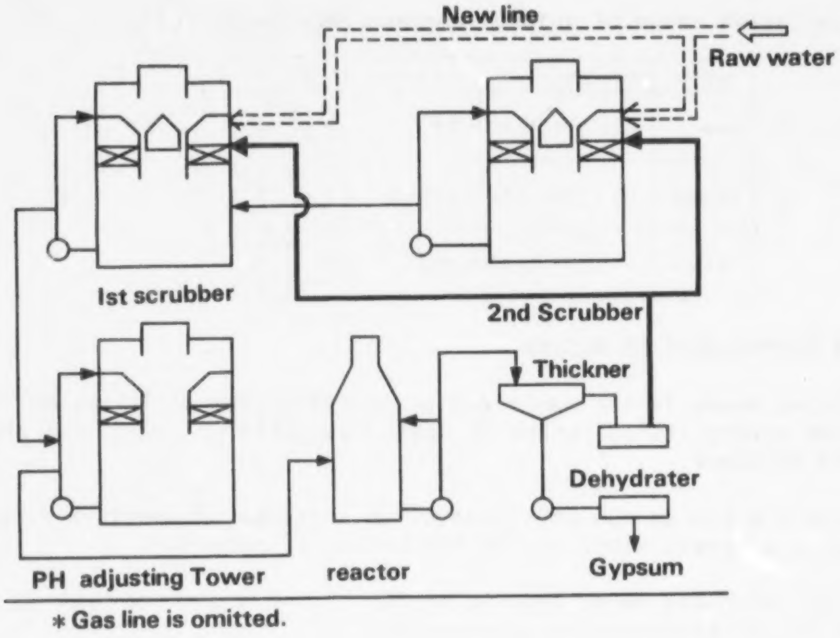

New line
Raw water
Ist scrubber
2nd Scrubber
Thickner
Dehydrater
Gypsum
PH adjusting Tower
reactor
* Gas line is omitted.

Figure 6. The Improvement of Eliminator Washing Water Line

As photo 4 and 5 shows, there is no sign of scale build up after approximately 4,500 hours since the start of supplying raw water to the washing water.

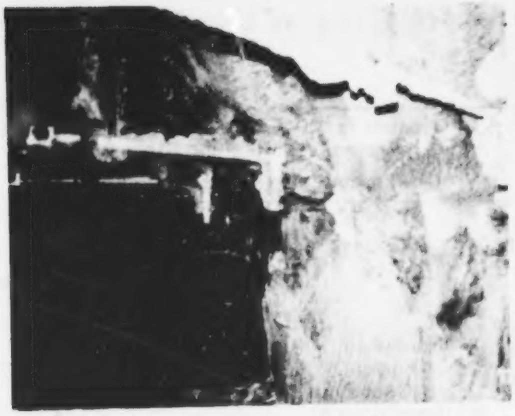


Photo 4 Before improvement

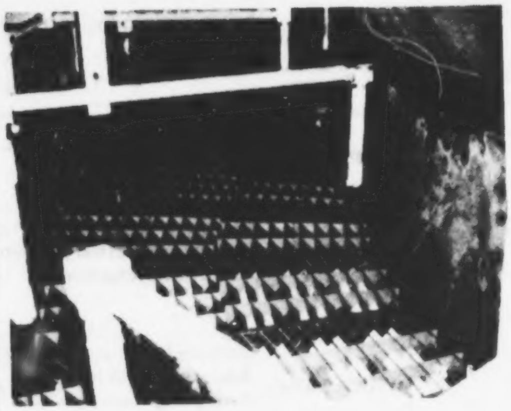


Photo 5 After improvement

9. MAJOR PROBLEMS OF FGD SYSTEM AT EACH PLANT

One problem of FGD system in EPDC's experience which has resulted in shutdowns of FGD system is caused by the scale build up in the scrubber when the SO_2 concentration in flue gas exceeds 1,000 ppm.

No trouble has been encountered in the FGD system at ISOGO Power Station where SO_2 concentration is very low.

Nevertheless, troubles related to scale build up have already been solved as previously mentioned.

The troubles encountered till the end of fiscal year of 1977 is summarized according to the number of troubles to have shut down FGD system and the description and the plant.

The shutdown time of FGD system means only the time when FGD system is out of service, meanwhile the plant is in service with low sulphur fuel, but does not include the case when a part of FGD system or one process line is out of service.

Table 8 shows major troubles of FGD system at each plant.

Table 8. Major Problems of FGD System at Each Plant

Plant	Absorption Process		Gypsum Product Process		Absorbent Receiving Process	
	Description	No.	Description	No.	Description	No.
ISOGO **(Stop Hour 0)**	**Failure of circulation Pump in Absorbing Tower (Poor Manufacture)**	**1**	**None**	**—**	**None**	**—**
TAKASAGO **(Stop Hour 544)**	**Scaling Build up in Absorbing Tower, PH Adjusting Tower (Stop Hour 498)** **BUF Trouble (Stop Hour 44)**	**#1 12** **#2 5** **#1 2**	**Trouble on Gypsum Conveyor (Stop Hour 5)**	**1**	**None**	**—**
TAKEHARA **(Stop Hour 0)**	**Trouble of Spray Nozzle in Cooling Tower (Half process Line stopped)**	**2**	**None**	**—**	**None**	**—**

10. FGD WASTE WATER TREATMENT SYSTEM

Japan has very severe waste water standards, therefore any waste water to be discharged from power stations must be treated to the extent it meets the standards.

The following items need to be controlled in the FGD waste water.

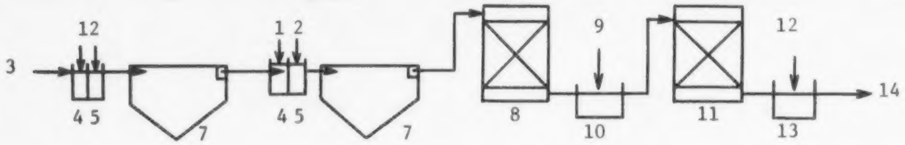
- o PH
- o SS
- o COD
- o Heavy Metal
- o F

(1) Description Water Treatment [1)]

In a wet FGD process, corrosion can occur because of the absorbed and concentrated chlorine ion (Cl^-) in the fuel. Accordingly, in order to maintain the chlorine ion (Cl^-) concentration below a certain level, blow down water in the process is required. Of primary importance in the blow down operation is the treatment of SS, fluorine ion (F^-), COD and some other heavy metal ions and the neutralization of discharge.

The SS, and some other heavy metals, are easily treated by using coagulative precipitation-filtering system; however, the required COD level in the FGD waste water (10 or 20 mg/ℓ) can not be achieved even by using activated charcoal absorption or oxidation with chlorine. It has been found, however, that the FGD waste water can be absorbed and removed by using a certain plastic absorbent. This process is actually in practical use at this time.

A water treatment system for FGD process is illustrated below:



1 Alkaline coagulant
2 Coagulant aids
3 Original water
4 PH adjustment tank
5 Caogulative tank
6 To dehydration equipment
7 Precipitation tank
8 Filter
9 Acid
10 PH adjustment tank
11 COD absorption bed
12 Alkali
13 Neutralization tank
14 Discharge

11. UTILIZATION OF BY-PRODUCT

Gypsum as a by-product is mainly taken by cement producing companies. Utilization of gypsum is indicated as follows and the market of gypsum is changeable in Japan.

- Tempering agent in cement
- Plaster
- Construction material as gypsum board

Table 9 shows annual product of EPDC's FGD system.

Certain type of catalyst is used to produce good gypsum in coal fired FGD process at Isogo and Takasago power stations so that size of gypsum crystal is large and the higher purity of gypsum is obtained compared to natural gypsum.

Table 10 shows the quality of gypsum to be unloaded from the power stations.

Table 9. Annual Gypsum Production of EPDC's FGD System

(Unit: metric ton)

Plant	1974	1975	1976	1977	Total
Isogo	-	-	30,755	36,608	67,363
Takasago	5,519	30,207	107,087	121,880	264,693
Takehara	-	-	7,815	66,890	74,705
Total	5,519	30,207	145,657	225,378	406,761

Table 10. The Quality of Gypsum from FGD

Composition	Wt (%)
$CaSO_4 \cdot 2H_2O$	96% or over
$CaSO_3 \cdot 1/2H_2O$	0.5% or less
$CaCO_3$	$0.5 \sim 1\%$
R_2O_3	1% or less
Residue insoluble by acid	1% or less

12. PROBLEM IN DISPOSAL OF GYPSUM

Table 11 shows FGD systems owned by power utility companies out of those in service in Japan at present.

The processes of FGD systems are almost of limestone-gypsum and the production capacity of gypsum from these FGD system goes up considerably. The market of the gypsum produced from FGD systems changes heavily according to the increase and decrease of supply and demand balance in the cement industry and the construction material industry.

EPDC asked in the past a cement company to take FGD gypsum with freight paid by EPDC when the demand of gypsum was quite low.

Problem will arise in future when new FGD systems are built because sales of gypsum itself may become impossible and even the disposition of gypsum shall be questionable.

Table 11. Desulfurization Installations in the Electric Utilities of Japan

Power Company	Desulfurization Process	Maker	Power Plant	Unit	Out Put (MW)	Fuel	Start - up	Gas Volume (Nm^3/H)	Capacity (%)	Efficiency (%)	Remarks
E P D C	Wet, Limestone-Gypsum	Mitsui Miike Machinery	Takasago	1	250	Coal	1975 - 2	840,000	100	93.3	EPDC: Electric Power Development Co., Ltd. IHI : Ishikawajima-Harima Heavy Industries Co., Ltd.
		Mitsui Miike Machinery	Takasago	2	250		1976 - 3	840,000	100	93.3	
		I H I	Isogo	1	265		1976 - 3	900,000	100	90	
		I H I	Isogo	2	265		1976 - 6	900,000	100	90	
		Babcock-Hitachi K.K.	Takehara	1	250		1977 - 2	852,000	100	94.2	
		I H I	Matsushima	1	500		1980 - 1	1,300,000	75	95	
		Babcock-Hitachi K.K.	Matsushima	2	500		1980 - 7	1,300,000	75	95	
Hokkaido	Wet, Limestone-Gypsum	Babcock-Hitachi K.K.	Date	1	350	H&C oil	1978 - 12	260,000	25	90	
		Babcock-Hitachi K.K.	Tomakomai-Shigashi	1	350	Coal	1980 - 8	610,000	50	90	
Tōhoku	Wet, Limestone-Gypsum	Mitsubishi Heavy Inds.	Hachinohe	4	250	Heavy Oil	1974 - 2	380,000	50	90	
		Mitsubishi Heavy Inds.	Higashi-Niigata	1	600		1976 - 6	420,000	25	90	
	Wet, Double Alkali-Gypsum	Kawasaki Heavy Inds.	Shin-Sendai	2	600		1974 - 3	420,000	25	96	tie up with Kureha Chemical Industries
	Wet, Sodium-Sulfuric Acid	Mitsubishi-Kakoki K.	Niigata	4	600		1977 - 3	760,000	50	90	Wellman-Load
	Wet, Double Alkali-Gypsum	Kawasaki Heavy Inds.	Akita				1977 - 9	1,050,000			
Tokyo	Dry, Active Carbon	Babcock-Hitachi K.K.	Kashima	3	600	Heavy Oil	1972 - 7	420,000	25	80	
	Wet, Limestone-Gypsum	Mitsubishi Heavy Inds.	Yokosuka	1	265		1974 - 1	400,000	50	90	
Chubu	Wet, Sodium-Sulfuric Acid	Mitsubishi-Kakoki K.	Nish-Nagoya	1	220	Heavy Oil	1973 - 5	620,000	100	90	
	Wet, Lime-Gypsum	Mitsubishi Heavy Inds.	Owase-Mita	1	375		1976 - 3	1,200,000	100	90	
		Mitsubishi Heavy Inds.	Owase-Mita	2	375		1976 - 5	1,200,000	100	90	
Hokuriku	Wet, Sulfuric Acid Dilution-Gypsum	Chiyoda Chemical Eng.	Toyama-Shinkō	1	500	Heavy & C. Oil	1974 - 10	750,000	50	90	
		& Construction Co., Ltd.	Fukui	1	350		1975 - 6	1,050,000	100	96	
Kansai	Wet, Lime-Gypsum	Mitsubishi Heavy Inds.	Amagasaki-Higashi	2	156	Heavy Oil	1972 - 3	100,000	25	90	
		Mitsubishi Heavy Inds.	Kainan	4	600		1973 - 12	400,000	25	90	
		Mitsubishi Heavy Inds.	Amagasaki-Higashi	2	156		1975 - 1	375,000	75	90	Additional Installation
	Wet, Limestone-Gypsum	Babcock-Hitachi K.K.	Osaka	3	156		1975 - 3	500,000	100	90	
		Babcock-Hitachi K.K.	Osaka	2	156		1975 - 12	500,000	100	90	
	Wet, Lime-Gypsum	Mitsubishi Heavy Inds.	Amagasaki-Higashi	1	156		1976 - 10	475,000	100	90	
	Wet, Limestone-Gypsum	Babcock-Hitachi K.K.	Osaka	4	156		1976 - 10	500,000	100	90	
Chūgoku	Wet, Limestone-Gypsum	Babcock-Hitachi K.K.	Mizushima	2	156	H. Oil	1974 - 4	310,000	66	80	
		Babcock-Hitachi K.K.	Tamashima	3	500	H&C Oil	1975 - 7	1,460,000	100	96	
		Babcock-Hitachi K.K.	Tamashima	2	350		1976 - 3	1,000,000	100	96	
		Mitsubishi Heavy Inds.	Shimonoseki	2	400	H. Oil	1977 - 4	1,200,000	100	90	
Shikoku	Wet, Double Alkali-Gypsum	Kawasaki Heavy Inds.	Anan	3	450	Heavy Oil	1975 - 8	1,260,000	100	97	tie up with Kureha Chemical Industries
		Kawasaki Heavy Inds.	Sakaide	3	450		1975 - 10	1,260,000	100	97	
Kyushu	Wet, Limestone-Gypsum	Mitsubishi Heavy Inds.	Karita	2	375	Heavy Oil	1974 - 6	550,000	50	90	
		Mitsubishi Heavy Inds.	Karatsu	2	375		1976 - 3	570,000	50	90	
		Mitsubishi Heavy Inds.	Ainoura	1	375		1976 - 4	730,000	75	90	
		Mitsubishi Heavy Inds.	Ainoura	2	500		1976 - 5	730,000	50	90	
		Mitsubishi Heavy Inds.	Karatus	3	500		1976 - 6	730,000	50	90	
	Wet, Double Alkali-Gypsum	Kawasaki Heavy Inds.	Buzen	1	500		1977 - 12	736,000	50	90	tie up with Kureha Chemical Industries
Others	Wet, Lime-Gypsum	Mitsubishi Heavy Inds.	Mizushima-Kyodo	5	156	1976 Heavy Oil	1976 - 1	611,000	100	90	
	Wet, Limestone-Gypsum	Mitsubishi Heavy Inds.	Niigata-Kyodo	1	350		1976 - 1	530,000	50	90	
		Mitsubishi Heavy Inds.	Niigata-Kyodo	2	350		1977 - 3	530,000	50	90	
		Mitsubishi Heavy Inds.	Sakata-Kyodo	1	350		1977 - 10	1,100,000	100	90	
		Mitsubishi Heavy Inds.	Sakata-Kyodo	2	350		1978 - 10	1,100,000	100	90	
		I H I	Sumitomo-Kyodo	3	156		1975 - 12	450,000	100	90	
		Mitsubishi Heavy Inds.	Fukui-Kyodo	1	250		1978 - 8	750,000	100	95	
		Mitsubishi Heavy Inds.	Kashima-Minami-Kyodo	2	-		1976 - 9	431,000	—	90	
	Wet, Sulfuric Acid Dilution Gypsum	Chiyoda Chemical E & C	Toyama-Kyodo	1	250		1975 - 9	750,000	100	92.5	

Following studies are under way as a solution of this problem.

- o Utilization of gypsum for reclamation material
- o Change to the more marketable material, i.e. elementary sulphur, as one of FGD by-products.

(1) Utilization of Gypsum for Reclamation Material

Tests were performed in 1976 by EPDC to determine if the gypsum can be used as reclamation material.

The tests have proved as follows.

- o No dissolution of heavy metal was observed from the gypsum, which was solidified by coagulator, to exceed the range of the waste water standards.
- o Gypsum itself cannot help being dissolved.

The research in this regard is now discontinued at present because any effective coagulator has been found to prevent dissolution of gypsum itself.

13. CONSTRUCTION PLAN OF NEW FGD SYSTEMS

EPDC is now constructing two 500MW coal fired power plants, and the plants are equipped with FGD systems which treat approx. 3/4 flue gas. Meanwhile EPDC is also planning one 700MW coal fired power plant with which a full capacity FGD system is scheduled to be equipped. Outline of new FGD systems is shown below.

(1) Matsushima Thermal Power Station

Matsushima Power Station is the one of coal fired plants to use imported coal. Generating capacity goes up 1,000MW consisting of two 500MW units.

No.1 plant will be operated commercially from January 1981 and No.2 plant from July 1981.

At the Power Station, EPDC will install wet limestone-gypsum FGD systems which have been proven at the other existing Power Stations of EPDC.

The capacity of the FGD systems is approx. 3/4 of the generating capacity of plants.

(a) Special features of Matsushima's FGD System

The special features of the FGD system for Matsushima Power Station are described below.

- o In order to burn the various kinds of imported coal, the prescrubber is installed to remove dust and impurities in flue gas before SO_2 removal.
- o The prescrubber and absorber are spray tower type which have very low pressure drop and high liquid-to-gas ratios.
- o In order to reduce consumption of energy and plant water, the reheating system is changed from after burning system to gas-gas heater of "Ljungstorm" type.
- o The process water treatment equipment is installed to reduce consumption of plant water, and to prevent scaling troubles.

(b) Design basis

The FGD system is designed based on our operating experiences in existing plants, and the results of research and development achieved by pilot plant tests.

The pilot plants for this system were installed at Isogo P/S and Takehara P/S. The tests were conducted for about one year.

The design basis of the FGD system for each 500MW power generating coal-fired boiler at Matsushima P/S is shown in Table 12.

(c) Process description

o Prescrubber

The flue gas is introduced from the two gas-gas heaters to a prescrubber through the two booster-fans, and is quenched to saturated temperature by contacting with cooling water.

At the same time, dust and impurities in flue gas such as HCl, HF, etc. removed in the prescrubber.

Removal of these impurities is very important in FGD process, so as to be described below.

- o to maintain high desulphurization efficiencies for the various gas conditions.
- o to improve the quality of gypsum.

o to reduce limestone stoichiometry.

o to prevent scaling problems, and corrosion of equipments.

Table 12. Outline of FGD System at Matsushima P/S

Item	Specifications
Gas flow to prescrubber	1,300,000 Nm^3/Hr (Wet) (Approx. 970,000 ACFM)
Flue gas temperature	Booster-fan outlet: 135°C Prescrubber inlet: Max. 110°C, Nor 85°C
Inlet gas SO_2 concentration	1,000 ppm Dry
Inlet gas particulate	300 mg/Nm^3 Dry (0.13 grain/SCFD)
Performance	
SO_2 Concentration	50 ppm Dry (Absorber outlet)
Particulate loading	30 mg/Nm^3 Dry (0.0125 grain/SCFD) (Absorber outlet)
Absorbent	98% $CaCO_3$, 5.5 Ton/Hr (Approx.) Size: 325 Mesh pass over 95%
By-product	Gypsum: 10.5 Ton/Hr (Approx.) Purity: Over 95% $CaSO_4$ $2H_2O$
Make-up water	Raw water: 49 Ton/Hr

The quenched gas is sent to the absorber through the vertical mist eliminator which is located in outlet duct of the prescrubber in order to prevent entrainment of mist accompanied in quenched gas.

- Absorber

The outlet gas from the prescrubber is led to a absorber, and the SO_2 in the flue gas is absorbed in the absorber by droplets of the limestone slurry which is sprayed from absorption spray nozzles.

At the top of absorber, a horizontal type mist eliminator is equipped.

The cleaned gas is fed to gas-gas heaters through this mist eliminator, where the entrained slurry droplets are separated.

- Reheating system

The desulfurized gas should be reheated for the purpose of protection of duct and stack corrosion, improvement of the atmospheric diffusion effect of flue gas from stack, and prevention of the white plume generation.

A "Ljungstom" type gas-gas heater is designed based on the results of research and development achieved by pilot scale tests at Takasago P/S.

- Gypsum production process

The slurry bled from the absorber includes calcium sulfites, calcium sulfates and limestone unreacted in the absorber.

The Calcium sulfites are oxidized to gypsum by air in the oxidation tower

The limestone unreacted is converted to gypsum by addition of sulfuric acid.

The gypsum is concentrated in the thickner, and then is dehydrated by the centrifuges.

- Process water treating system

A part of process water is bled from the thickner to keep water balance of process, because the fresh water is supplied to the mist-eliminator in the absorber.

The process water treating equipment is equipped for reusing.

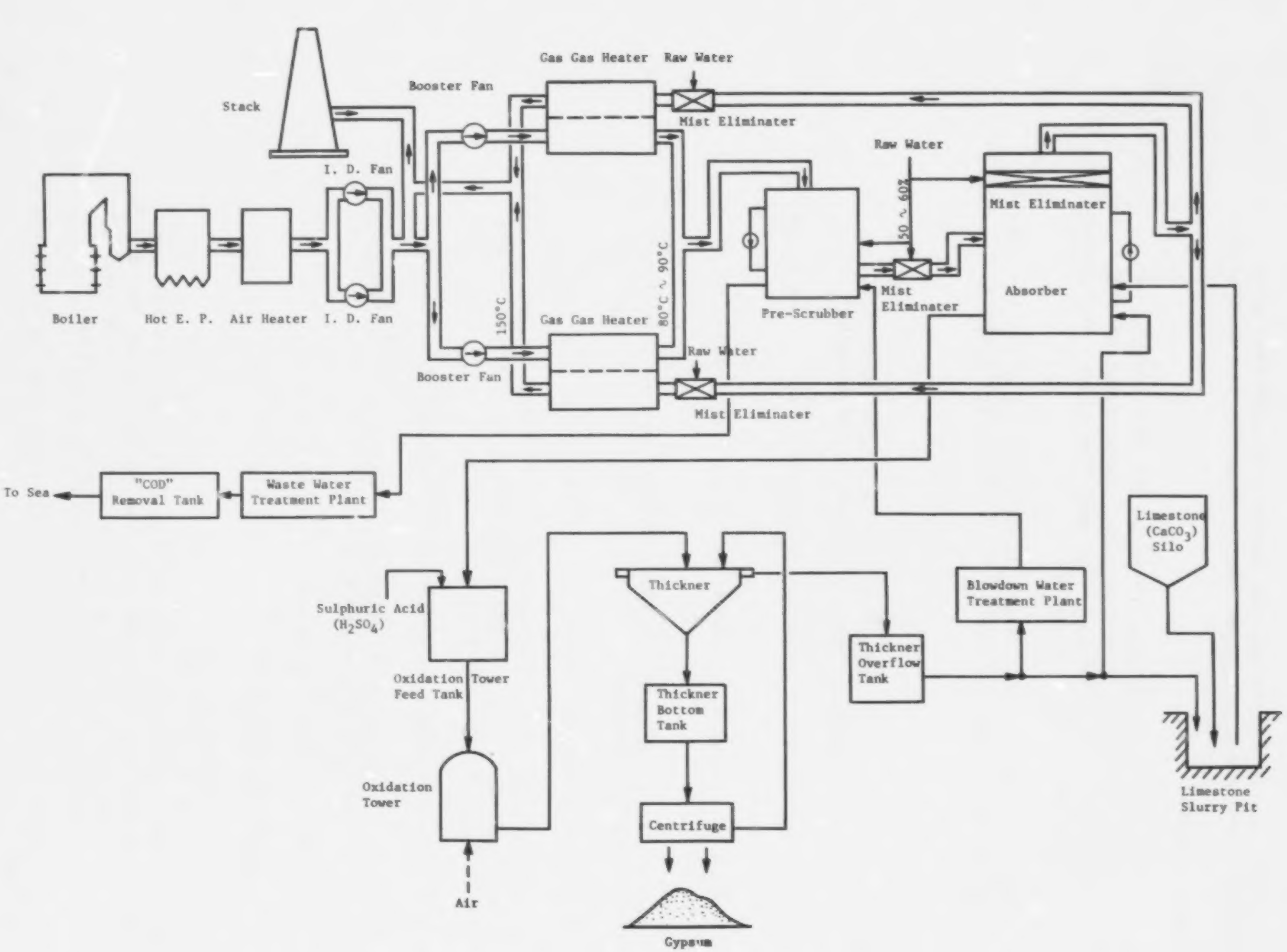
The SS and Ca^+ in the bleed water are treated by this equipment and then is supplied to the prescrubber.

Water quality (outlet)	100 ppm (as Ca^+)
Water quality	15 m^3/Hr

- Waste water treating system

A part of the prescrubber recycle liquor is bled from the prescrubber to keep the dissolved chlorine concentration in recycle liquor at 5,000 ppm or less, and to discharge the dust removed, which is preferable for the materials of construction.

The waste water is treated in this system and then is discharge to the sea.



Stack
Booster Fan
Gas Gas Heater
Raw Water
Mist Eliminater
I. D. Fan
Raw Water
50 ~ 60%
Mist Eliminater
Boiler
Hot E. P.
Air Heater
I. D. Fan
150°C
80°C ~ 90°C
Pre-Scrubber
Mist Eliminater
Absorber
Gas Gas Heater
Booster Fan
Raw Water
Mist Eliminater
To Sea
"COD" Removal Tank
Waste Water Treatment Plant
Limestone (CaCO3) Silo
Thickner
Blowdown Water Treatment Plant
Sulphuric Acid (H2SO4)
Oxidation Tower Feed Tank
Thickner Bottom Tank
Thickner Overflow Tank
Oxidation Tower
Centrifuge
Limestone Slurry Pit
Air
Gypsum

Figure 7. The Flow Chart of Matsushima's FGD

(2) Takehara Thermal Power Station

EPDC is planning to extend No.3 plant at Takehara Power Station. Table 13 is the outline of extension plant but the specification has not been finalized.

Table 13. Takehara No.3 Plant Extension Plan

Generating capacity	700 MW
Fuel	Coal
FGD capacity	2,700,000 Nm^3/H (Full capacity)
Process	Limestone-gypsum
Commercial operation	July, 1982

Specific features of the FGD system design for No.3 plant are summarized as Table 14.

Table 14. Specific Features of FGD System for Takehara No.3 Plant

- o Treatment of Nitrogen Removal from FGD Waste Water
- o New Position of DeSOx Fan

(a) Treatment of nitrogen removal from FGD waste water

As DeNOx system is installed for No.3 plant, ammonia leaked from DeNOx system goes into FGD system where it is dissolved in the process water and discharged in the FGD Waste Water after concentrated.

Biological treatment is introduced to remove nitrogen ion such as NO_3^- contained in the FGD waste water.

The success of developing biological treatment system is due to the effort of some Japanese manufacturers which have been studying jointly with EPDC the removal of nitrogen from the FGD waste water.

(b) New position of DeSOx fan

Gas-gas heater (called as GGH) will be applied like the FGD system of Matsushima Power Station, however, the severer SOx regulation will be imposed to No.3 plant of Takehara Power Station by a local autonomous body and any boost up fan cannot be installed in series of ID Fan like that of Matsushima as shown in Figure 8, because the leakage of flue gas in GGH will cause a drop of total desulphurization efficiency.

With this reason, if the position of the fan is moved to the one between GGH and FGD system (called as $DeSO_x$ Fan) the leakage through GGH will occur from the treated gas side to untreated gas side, then the desulphurization efficiency of the FGD system becomes the same value of total desulphurization efficiency of the plant.

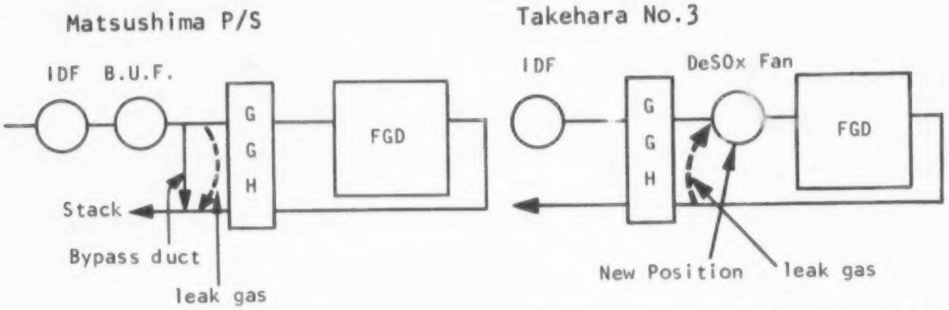


Figure 8. Location of DeSOx Fan

14. STUDY OF NEW FGD PROCESS

It is indispensable to take full countermeasures for flue gas treatment of coal fired power stations in Japan along with enforcement of severer popullation regulations.

EPDC has achieved a good prospect of commercialization of NOx Removal System through the past researches and developments.

EPDC's papers were presented in this regard at DeNOx Seminar in Denver Colorado in October 1978 sponsored by EPRI.

Problems are expected to arise on the FGD Systems installed at coal fired power stations because the waste water treatment system will become complicated with an introduction of ammonia treatment system in conjunction with NOx removal system.

The other problems may arise in future on the disposal of FGD gypsum which will be produced excessively in the market.

EPDC is challenging further studies on new FGD systems bearing the backgrounds in mind.

(1) Studies of New FGD System

New Dry FGD has recently been developped. The process applies activated carbon with ammonia injection which evaluated as superior as to wet process in terms of performance capability and operation economy.

The dry process will prove better evaluation for EPDC with a combination with NOx Removal System.

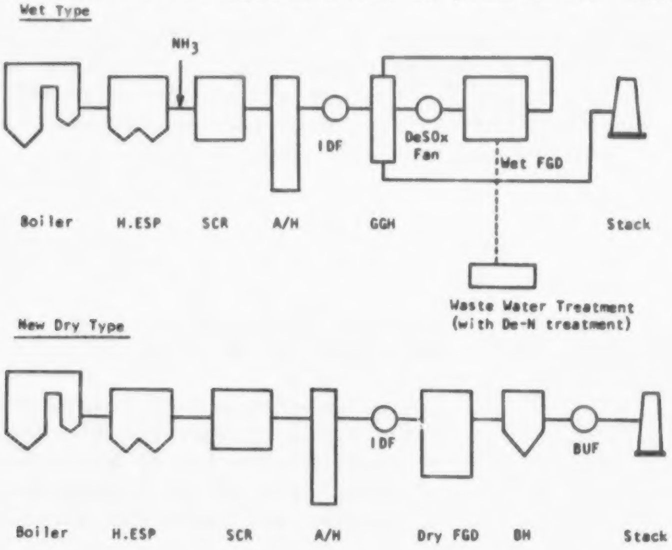
Table 15 shows the advantage of the new dry process.

Table 15. Advantage of The New Dry Process

- o Simpler Flue Gas Treatment System at a coal fired power plant
- o Elemental Sulphur as a by-product
- o Simpler Waste Water Treatment System

(2) New FGD System

Figure 9 shows the comparison of flow diagrams for New FGD system and the FGD system and the FGD system which is planned for Takehara No.3 Plant.


Wet Type
NH3
IDF
DeSOx Fan
Wet FGD
Boiler
H.ESP
SCR
A/H
GGH
Stack
Waste Water Treatment
(with De-N treatment)
New Dry Type
IDF
BUF
Boiler
H.ESP
SCR
A/H
Dry FGD
BH
Stack

H.ESP: High Temperature Electrostatic Pricipitator A/H: Air Preheater
SCR: Selective Catalytic Reduction System GGH: Gas/Gas Exchanger
IDF: Induced Draft Fan BUF: Boost-up Fan BH: Bas House

Figure 9. New FGD System and FGD System for Takehara No.3

EPDC is conducting test of 10,000 Nm3/H at a EPDC's power station starting the test from November 1978.

EPDC will present details after observing tests status and others.

SUMMARY

(1) EPDC has judged Wet Limestone-gypsum process as the best process.

(2) Basically no problems have been encountered on Wet Limestone-gypsum process. The FDG systems of its process have been in service satisfactorily and its availability is the same as the boiler availability.

(3) Desulphurization efficiency changes slightly according to boiler load variation but at least more than 85% of the efficiency is maintained in average.

(4) GGH and Denitrogen Treatment is applied, as new peripheral technology, to FGD systems at newly built power stations.

(5) New FGD is under study in connection with the introduction of NOx Removal Systems and its researches and developments are underway.

ACKNOWLEDGEMENT

The fact that FGD systems have been in service with high reliability and good performance is mainly dependent upon special attention and maintenance of the engineers and the operators working at EPDC's Power Stations as well as the efforts exerted by the manufacturers which have been making joint development with EPDC.

REFERENCE:

1) T. Hayase, K. Mouri ; "State of Development and use of the Desulphurization in Japan"

Symposium of state of development and proving in operational tests of desulphurization of Waste Gases of Power Stations by Czechoslovac Scientific and Technical Society.

March 1978

CURRENT ALTERNATIVES FOR FLUE GAS DESULFURIZATION (FGD) WASTE DISPOSAL - AN ASSESSMENT

By

Chakra J. Santhanam, Richard R. Lunt and Charles B. Cooper

Arthur D. Little, Inc.
Cambridge, Mass. 02140

ABSTRACT

With increasing coal utilization in industrial and utility boilers, generation of coal ash (fly ash and bottom ash), and flue gas desulfurization (FGD) wastes, which together comprise flue gas cleaning (FGC) wastes, is expected to increase dramatically over the next twenty (20) years. Since most of the FGC wastes generated will be disposed of, rather than utilized, these wastes represent significant potential sources of environmental pollution unless proper disposal technology is employed. Continuing research and development efforts sponsored by EPA, EPRI, and others in recent years have provided substantial information on environmentally sound disposal techniques.

This paper discusses the current state-of-the-art of flue gas desulfurization (FGD) waste disposal with focus on wastes from nonrecovery systems. The paper includes a review of the following areas:

- Production and categorization of FGC wastes;
- Disposal options presently in use and/or with future potential; and
- Environmental issues associated with various disposal options.

1.0 INTRCDUCTION

As coal utilization in utilities and large industrial boilers increases, the quantity of flue gas cleaning (FGC) wastes, particularly those associated with flue gas desulfurization, will increase dramatically. The preponderant part of these FGC wastes will be discharged disposal. Over the long term, utilization is expected to grow but at a slower rate than that of FGC waste generation. Table 1 shows projections of coal ash and flue gas desulfurization (FGD) wastes through 2000.

In the past, utilities operating FGC systems have typically disposed of wastes by storage in ponds, often without provision for control of overflows or seepage into groundwater. However, several factors will dramatically influence disposal options in the coming years.

a. An increase in coal-fired capacity in the United States. In 1976 the total U.S. coal-fired electric utility generating capacity was estimated at over 191,000 MW in 399 plants (3). The estimated capacity is expected to increase by 1986 to over 326,000 MW (4). Use of coal in large industrial boilers (+25 MW equivalent or larger) is likely to further increase the total coal-fired capacity.(1)

b. A major increase in the application of scrubber technology by utilities and a consequent increase in FGD waste generation. At present over 16,000 MW of generating capacity at some thirty plants utilize FGD systems. As of September 1978, over 59,000 MW of capacity have been committed (5). Future increases are likely to be even more dramatic.

c. Advances in stabilization technology for FGD wastes which permit landfill disposal of partially dewatered solids instead of ponding of difficult to handle sludges. In the future, disposal of wastes in managed fills is likely to be encouraged. In many cases this will require stabilization prior to disposal.

d. Regulatory developments including the Clean Air Act of 1977 and the Resource Conservation & Recovery Act of 1976 (RCRA). New Source Performance Standards (NSPS) for criteria pollutants are now under review by the EPA and may be significantly tightened. Similarly, the recent issuance of proposed guidelines provides impetus to environmentally sound disposal of FGC wastes.

R&D efforts and regulatory development have focused on characterization and categorization of FGC wastes, determination of the mechanisms by which pollutants may enter the environment from a disposal site, assessment of the environmental impacts and development of disposal criteria and guidelines.

Table 1

Projected Generation of Coal Ash and FGD Wastes

			PROJECTED			
	1975		1985		2000	
	10^3 Metric Tons	% of Total	10^3 Metric Tons	% of Total	10^3 Metric Tons	% of Total
Coal Ash						
Industrial	-	-	8,590	12	19,950	19
Utility	-	-	64,440	88	84,800	81
Total	52,060	-	73,030	100	104,750	100
FGD Wastes						
Industrial	-	-	1,090	5	5,260	15
Utility	-	-	21,050	95	29,860	85
Total	6,200	-	22,140	100	35,120	100

Source: (1,2)

2.0 GENERATION OF FGD WASTES

2.1 Overview on FGC Technology

Ash Collection Technology

Coal-fired utility and industrial boilers generate two types of coal ash--fly ash and bottom ash. (Economizer ash and mill rejects are lumped into the two major categories here.) Both constitute the non-combustible (mineral) fraction of the coal and the unburned residuals. Fly ash, which accounts for the majority of the ash generated, is the fine ash fraction carried out of the boiler in the flue gas. Bottom ash is that material which drops to the bottom of the boiler and is collected either as boiler slag or dry bottom ash, depending upon the type of boiler.

The total amount of coal ash produced is directly a function of the ash content of the coal fired. Thus, the total quantity of ash produced can range from a few percent of the weight of the coal fired to as much as 35%. The partitioning of ash between fly ash and bottom ash ususally depends upon the type of boiler. Standard pulverized coal-fired boilers typically produce 80-90% of the ash as fly ash. In cyclone-fired boilers, which are frequently used to burn lignite, the fly ash fraction is usually somewhat less; in some cases bottom ash constitutes the majority of the total ash created.

Collection of bottom ash (or boiler slag) does not involve systems outside the boiler itself. The key technology issue is the handling of bottom ash. Fly ash, however, is a major source of particulate emissions and with regulatory requirements has required major collection systems. Control of particulate emissions from pulverized-coal-fired steam generators is rapidly becoming a significant factor in the siting and public acceptability of coal-burning power plants. The particulate emissions limit under current NSPS set by the EPA for large, new coal-fired boilers is 0.043 grams/10^6 joules (0.1 lb/10^6 Btu). Some states have requirements more restrictive than this. Furthermore, the NSPS are now under review and are expected to be tightened significantly.

Fly ash carried in the flue gas stream can be collected in a number of ways to meet the current particulate emission control limitations as noted above. Typical methods historically employed include mechanical collection, electrostatic precipitation, fabric filtration and wet scrubbing. However, the tightening regulatory requirements support two criteria for future fly ash collection systems:

- The collector must be efficient in removing sub-micron particulate matter. This criterion eliminates all mechanical collectors and many wet scrubber systems from consideration as the only systems. Mechanical collectors may, however, function as a first unit followed by a more efficient collector.

- The collector must be available commercially and be proven in a utility boiler application. This constraint eliminates, for the immediate future, many hybrid wet scrubber systems and novel collectors that are now under development. In the long run, however, it is conceivable that such advanced systems may be used at least in some instances.

It appears that electrostatic precipitators and fabric filters will be the only systems capable of meeting the requirements in the foreseeable future. In addition, developments in effluent guideline standards strongly point to dry fly ash handling systems in new plants (at present, the major portion of ash is handled by wet sluicing where dry collectors are employed).

FGD Technology

The implementation of flue gas desulfurization (FGD) technology for the control of SO_2 emissions from the combustion of fossil fuels in industrial and utility boilers is rapidly growing in the United States. At present, FGD systems are in operation on over 16,000 megawatts of utility generating capacity at some 30 different plants throughout the country, and more than 40 industrial steam plants are equipped with FGD systems. By the end of 1979, the total capacity of FGD systems in operation on utility and industrial boilers is expected to exceed 25,000 megawatts (equivalent). The degree of SO_2 control ranges from less than 50% SO_2 removal efficiency to over 90%, depending upon the type of FGD systems, the sulfur content of the fuel, and the applicable SO_2 emission regulations.

The growth in FGD systems on fossil-fuel-fired boilers in the United States over the next 20 years will be principally dependent upon the growth in utility and industrial boiler capacity, current and future SO_2 emission regulations, and the impact of alternative desulfurization approaches to current and developing FGD technology. An important factor may be the use of existing and enhanced coal-cleaning techniques.

A wide variety of FGD processes have been developed for application on utility and industrial boilers. In general, the technology can be grouped into two categories: nonrecovery, or throwaway systems, which produce a waste material for disposal; and recovery systems, which produce a saleable byproduct (either sulfur or sulfuric acid) from the recovered SO_2. Nonrecovery processes make up the overwhelming majority of the technology. Nine different processes and process variations can be considered to be commercially available, seven of which are nonrecovery systems. These seven processes constitute more than 95% of the capacity currently in operation on utility and industrial boilers, a trend which is expected to continue for the foreseeable future. Table 2 summarizes the applications of FGD process technologies for systems expected to be in operation by the end of 1979.

Table 2

Summary of FGD Systems Expected To Be In Commercial Operation on Utility and Industrial Boilers in 1979

	Utility		Industrial	
	No. of Plants	Capacity (MW)	No. of Plants	Capacity (10^3scfm)
Nonrecovery				
Direct Limestone-Conventional Forced Oxidation	19	11,780	1	50
Direct Lime	13	7,305	1	85
Alkaline Ash	2	1,170	0	--
Dual Alkali	3	1,105	8	1,082
Once-Through Sodium	1	510	26	4,954
Ammonia	0	--	7	552
Total	38	21,870	43	6,722 (~3000 MW-eq)
Recovery				
Wellman-Lord	2	735	0	--
Citrate	0	--	1	104
Mag-Ox[a]	0	--	0	--
Total	2	735	1	104 (~50 MW-eq)

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[a] Two systems have been commercially operated on utility coal-fired boilers but these are not currently in operation.

Source: Arthur D. Little, Inc.

Nonrecovery Systems: Nonrecovery processes in general can be subdivided into two groups, wet processes and dry processes. Wet processes involve contacting the flue gas with aqueous slurries or solutions of absorbents and produce wastes in the form of solutions or slurries for direct discharge or further processing prior to disposal. In some cases, waste slurries are partially dewatered and further processed to produce a soil-like material for landfill. Dry processes, on the other hand, produce essentially moisture-free solids through dry injection of absorbents into the flue gas or the use of spray dryers. All nonrecovery processes now in operation as well as those due to come on line in 1979 involve wet scrubbing. However, a number of contracts have been signed for the application of dry systems to utility boilers which will start up in the early 1980's.

Of the seven different types of nonrecovery processes now in commercial operation on industrial and utility boilers, five involve conversion of the SO_2 to some form of solid waste (sludge) for disposal in either wet ponds or landfills:

- Conventional direct lime scrubbing,
- Conventional direct limestone scrubbing,
- Limestone scrubbing with forced oxidation,
- Alkaline fly ash scrubbing, and
- Dual alkali.

Two systems produce a soluble waste which is discharged as an aqueous liquor to holding ponds or wastewater treatment systems:

- Once-through sodium scrubbing, and
- Ammonia water scrubbing.

As shown in Table 2, essentially all utility applications of nonrecovery technology involve solid waste-producing systems. In contrast, a large majority of industrial boiler applications of FGD involve the production of liquid wastes.

Both types of wet scrubbing nonrecovery systems can usually withstand relatively high levels of particulate, and many in the past have been designed for simultaneous SO_2 and particulate removal. Approximately 40% of FGD systems currently in operation on utility boilers and about 80% of those in operation on industrial boilers serve as combined particulate and SO_2 control systems. However, most systems being installed today on utility-scale boilers follow high efficiency electrostatic precipitators in order to ensure reliable service of the FGD system.

Dry nonrecovery processes have not yet been commercially demonstrated in the United States, although at least three systems for full-scale utility applications are in the early stages of planning or design.

Three different approaches to dry scrubbing for producing solid wastes have been actively pursued (6):

- Injection of solid sorbents into the flue gas stream with collection of sorbents downstream in a particulate control device;
- Injection of solid sorbents into the boiler combustion zone; and
- Contacting of flue gas with alkali sorbent slurries in a spray dryer.

All of these approaches involve simultaneous particulate and SO_2 control, and all offer the advantage of not requiring flue gas reheat, which wet processes generally do require.

Recovery Processes: As in the case of nonrecovery processes, recovery processes can also be categorized into wet and dry according to the mode of SO_2 removal. They can be further classified according to the type of byproduct produced: concentrated SO_2 for conversion to sulfur or sulfuric acid; sulfur only; or acid only.

At present, only two process technologies have been commercially demonstrated on large industrial - or utility-scale boilers--the Wellman-Lord process and magnesium oxide scrubbing. Another, the citrate scrubbing process, is currently being commercially tested on a large industrial boiler. All three of these are wet scrubbing processes.

The total capacity attributable to these three technologies (including magnesium oxide system not now in operation) is less than 5% of the total FGD operating capacity in 1979; and market share is expected to remain below 5% of the total installed FGD capacity on boilers in the United States through the mid-1980's. Since recovery systems represent such a small fraction of the market, and produce only a small amount of wastes in comparison to nonrecovery systems, discussions throughout the remainder of this paper will focus on wastes from nonrecovery systems.

2.2 General Composition and Categorization of Nonrecovery FGD Wastes

Major Components

The quantity and characteristics of FGD wastes produced from a combustion system depend on a variety of factors including:

- Composition of the coal (ash and sulfur content);
- Type of combustion (boiler) system and its operating conditions;
- Type of particulate collection system and its operating conditions;
- Type of FGD system and its operating conditions; and
- Degree of SO_2 control required.

The principal substances making up the solid phase of FGD wastes are calcium-sulfur salts (calcium sulfite and/or calcium sulfate) along with varying amounts of calcium carbonate, unreacted lime, inerts, and fly ash. The ratio of calcium sulfite to calcium sulfate is a key parameter (the latter, usually present as $CaSO_4 \cdot 1/2\ H_2O$ or as gypsum, $CaSO_4 \cdot 2H_2O$) and will depend principally upon the extent to which oxidation occurs within the system. Oxidation (and thereby sulfate content) is generally highest in systems installed on boilers burning low sulfur coal or in systems where oxidation is intentionally promoted. When the sulfate content of the waste solids is low, calcium sulfate can exist with calcium sulfite as a solid solution of hemihydrate crystals ($CaSO_x \cdot 1/2\ H_2O$). Data from pilot plant, prototype, and full-scale FGC system operations indicate that up to about 25% of the total calcium-sulfur salts can be present as $CaSO_4 \cdot 1/2\ H_2O$ in solid solution with $CaSO_3 \cdot 1/2\ H_2O$. At higher calcium sulfate levels, gypsum ($CaSO_4 \cdot 2H_2O$) becomes the predominant form of calcium sulfate. It is expected that at very high levels of oxidation (greater than 90% oxidation of the SO_2 removed) calcium sulfite can also form a solid solution with gypsum ($CaSO_x \cdot 2H_2O$) analogous to the solid solution of hemihydrate salts formed at low sulfate levels.

Because the differences in the crystalline morphology of hemihydrate and dihydrate solids not only reflect the chemical composition but also to a large extent dictate the physical and engineering properties of FGC wastes, it is convenient to classify FGC wastes on the basis of the calcium sulfate content. Three such categories have been selected, as follows:

Category	Predominant Crystalline Form
Sulfate-rich ($CaSO_4/CaSO_x \leq 0.90$)	Dihydrate
Mixed ($0.25 \geq CaSO_4/CaSO_x \leq 0.90$)	Dihydrate and hemihydrate
Sulfite-rich ($CaSO_4/CaSO_x \leq 0.25$)	Hemihydrate

where $CaSO_x$ is the total calcium-sulfur salt content. This categorization will be employed in the ensuing discussions throughout this paper.

Factors which tend to influence the amount of sulfite in FGC wastes (i.e., the extent of oxidation) include: boiler excess air, type of scrubber, use of forced oxidation, presence of oxidation inhibitors or catalysts in fly ash, reagents (or water makeup), type of reagent, pH in the scrubber loop, sulfur content of the coal and the degree of SO_2 removal.

In general, it is possible to relate the three general categories of wastes indicated above and their associated crystalline morphologies with various types of FGC process technologies and their applications according to the coal sulfur content. Such a matrix relationship is shown in Table 3. As indicated, dual alkali and conventional direct lime scrubbing systems using either carbide or Thiosorbic lime almost exclusively produce sulfite-rich wastes. Such systems are generally applied to medium and high sulfur coal-fired boilers, and attempts are made to minimize oxidation. On the other hand, alkaline ash and limestone

Table 3

Matrix of Untreated FGD Wastes Generation-
Nonrecovery Solid Waste Producing Systems

No.	Waste Type[a]	Chystalline Morphology	Low/Med. Sulfur Coal[b]				High Sulfur Coal[c]			
			DL[d]/AA[e]	DLS[f]	LSFO[g]	DA[h]	DL	DLS	LSFO	DA
1.	Sulfite-Rich ($CaSO_x \cdot 1/2\ H_2O$)	Needles	-	-	-	-	-	-	-	-
		Platelets	-	-	-	-	✓	✓	-	-
		Agglomerates	-	-	-	-	✓	?	-	✓
2.	Mixed Sulfite/Sulfate ($CaSO_x \cdot 1/2\ H_2O + CaSO_4 \cdot 2H_2O$)	Needles or Platelets + Platy[i]	✓	✓	-	✓	✓	✓	-	✓
3.	Sulfate-Rich ($CaSO_4 \cdot 1/2\ H_2O$ +/or $CaSO_4 \cdot 2H_2O$)	Needles or Platy[i]	✓	✓	✓	✓	?	✓	✓	-

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[a] Sulfite-Rich = $CaSO_4/CaSO_x$ $x \leq .25$

Sulfate-Rich = $CaSO_4/CaSO_x$ $x \geq .9$

[b] Low/Med. Sulfur Coal $\leq$ 2% S

[c] High Sulfur Coal >2% S

[d] Conventional Direct Lime Process

[e] Alkaline Ash Scrubbing

[f] Conventional Direct Limestone Process

[g] Limestone with Forced Oxidation

[h] Dual Alkali Process

[i] Resembling rhombohedral cleavage fragments

Notes:

✓ Refers to the particular waste type as the common waste product from the type of coal and process.

? Some question on this.

forced oxidation systems produce sulfate-rich wastes almost exclusively. And conventional direct lime (using commercial lime) and limestone systems can produce either sulfite-rich, sulfate-rich, or mixed wastes depending upon the sulfur content of the coal and the manner in which the scrubber systems are operated.

Fly ash will be a principal constituent of FGD wastes only if the scrubber serves as a particulate control device in addition to SO_2 removal, or if fly ash separately collected is admixed with the wastes from SO_2 scrubbing. The amount of fly ash, therefore, can range from nil to as much as 80% of the total dry weight of the wastes produced. More than 85% of the total weight of fly ash is made up of silica, alumina, and iron, calcium and magnesium oxides. These will appear in the wastes to the extent that fly ash is present.

Minor and Trace Components

FGD waste solids from wet scrubbing processes carry with them occluded liquor which contains dissolved solids. The amount of liquor is a function of the degree of dewatering prior to discharge. The major soluble ions usually present include calcium, chloride, magnesium, potassium, sodium, sulfite and sulfate. Together, these can amount to as much as 10-15% of the two solids (dry basis).

A variety of trace elements are also present in FGD wastes and derive from a number of sources: coal, where they are present either in mineral impurities or as organometallic compounds; makeup chemicals for the FGD system; and FGD process makeup water. The principal source of trace elements, though, is from the coal; and the levels of trace elements depend primarily on their level in the coal, the amount, if any, of ash that is collected or admixed with the wastes, and the efficiency of the scrubber system in capturing trace metal vapors and fine particulate. Since most of the elements in coal are not highly volatile and will be retained in the ash matrix (either as fly ash or bottom ash), the presence and concentrations of most trace elements will depend upon whether fly ash is simultaneously removed with SO_2 or admixed with the waste calcium-sulfur salts. The concentrations in the wastes of those elements that are most highly volatile (notably arsenic, mercury, selenium, beryllium, chloride, and fluoride) will be a function of the extent to which they are present and released from the coal and, more importantly, the efficiency with which they are captured in the scrubber.

2.3 Dewatering of FGC Wastes

Most unthickened slurry wastes produced by FGC systems contain on the order of 5-15 wt% suspended solids. In order to avoid the unnecessary discharge of large amounts of process liquor, these wastes are frequently mechanically dewatered prior to being discharged from the process. Primary dewatering is usually accomplished using thickener/clarifiers or settling ponds. Primary dewatering is virtually universally practiced in order to reduce sludge volume and conserve water. Secondary methods of dewatering are also sometimes employed. These include vacuum filtration and centrifugation. Secondary dewatering is only employed as a precursor to dry impoundment in order to improve the handling properties of the wastes prior to truck transport or stabilization. In general the dewaterability of FGD wastes varies with the sulfite/sulfate content of the waste and the amount of fly ash present. Sulfite-rich wastes can typically be dewatered to 40-65 wt% solids, while sulfate-rich wastes can usually be filtered to 65-85% solids. The presence of fly ash and unreacted limestone can often improve the dewaterability, but significant improvements usually only occur for wastes with poor dewatering characteristics (within each general category of waste type).

Table 4 summarizes dewatering practices for full-scale FGC systems in operation on utility boilers as of November 1978 and shows some interesting trends in dewatering practices in the utility industry:

- No simultaneous SO_2 and fly ash control systems or wet particulate scrubbing systems employ secondary methods of dewatering (i.e., filtration or centrifugation) for FGC waste dewatering, although a number of the plants do dispose of wastes via dry impoundment of wastes reclaimed from secondary settling ponds.
- The overwhelming majority of the FGD capacity for SO_2 removal only involves thickening and filtration or centrifugation for dry impoundment of the wastes. This trend is expected to continue for the foreseeable future. About 6,700 megawatts of new, nonrecovery FGC capacity producing solid wastes are expected to be on-line in 1979, all of which will be devoted to SO_2 control only. Of this total, approximately 85% will utilize some form of dry impoundment for waste disposal, and more than two-thirds of these will employ either filtration or centrifugation for waste dewatering.

2.4 Stabilization Processes

There are now more than two dozen processes for solidification/stabilization of many types of sludges and difficult to handle wastes. The state of development of these processes ranges from laboratory-scale on a number of different types of wastes.

There are basically three methods by which stabilization processes can improve the disposability of wastes.

- First, through improvements in the physical characteristics of the wastes to the extent that they are more easily handled. This frequently leads to better control/management of the disposal area, resulting in reduced impacts relating to physical stability and contamination of ground and surface waters.
- Second, through decrease in the exposure of the wastes by reducing surface area and/or permeability or by encapsulating the wastes, thus limiting the contact of groundwater (or infiltration water) with the waste.

Table 4

Summary of FGC Waste Dewatering Practices for Operating Utility Scrubbers[a]

	Dewatering Practices Employed				
	Pond Settling[d]	Thickening	Thickening/ Pond Settling	Thickening/ Filtration	Thickening/ Centrifugation
		(No. of Plants/Total Capacity, MW)			
Scrubber System Mode					
• SO_2 Removal					
- Low Sulfur Coal[b]	3/1570	2/365	--	3/1045	1/1585
- High Sulfur Coal[c]	1/550	3/185	--	6/2515[e]	--
• SO_2 + Ash Removal					
- Low Sulfur Coal[b]	2/1085	3/2185	--	--	--
- High Sulfur Coal[c]	2/885	1/1650	--	--	--
• Wet Particulate Removal					
- Low Sulfur Coal[b]	7/1220	3/865	2/1175	--	--
Total	15/5310	12/6250	2/1175	9/3560	1/1585

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[a] Basis: November 1978

[b] Generally $\leq$ 1.5% sulfur

[c] Generally > 1.5% sulfur

[d] In addition to dewatering, settling pond acts as final disposal site in 10/3330 of those indicated.

[e] Includes two plants (totaling 920 MW) whose scrubber systems remove ash but have ESP's for primary ash removal.

- Finally, by chemical reaction with the waste, limiting the solubility of chemical constituents that would otherwise be readily accessible either through flushing of interstitial liquor or solubilization.

Different techniques usually emphasize one or two of these factors. The applicability and "success" of a particular process, therefore, will depend importantly upon the chemical and physical properties of the waste, the disposal site characteristics, and the waste-handling constraints.

At the risk of oversimplification, most all stabilization processes generally can be categorized into one of about six groups, according to the manner in which the wastes are treated. Table 5 lists the principal processes of each type, indicating the vendors and status of the process.

Stabilization of FGD Wastes

A number of the processes listed in Table 5 have been tested on FGD wastes, mostly in the bench scale. Justification of the use of additives to improve the physical characteristics of FGD wastes has been based on improvement in strength, reduction in compressibility, and reduction in permeability caused by an increase in solids content or the formation of permanent bonds between particles. The additives most advantageous then would be those available at low cost in large quantities (e.g., fly ash) and those effective as cementing agents (e.g., Portland cement). Combinations of additives may produce both types of improvement (e.g., fly ash plus lime). A limited amount of study has been devoted to the evaluation of simple additives such as fly ash, lime, and Portland cement. These studies are discussed later.

At present, there are two approaches which have achieved commercial applicability for calcium-based FGD wastes: addition of lime and fly ash for dry impoundment systems (currently marketed by IU Conversion Systems, Inc., and others) and the proprietary technology developed by the Dravo Corporation involving the use of processed blast furnace slag as the additive for stabilization in wet ponds. Other additives and stabilization approaches for calcium-based wastes have been laboratory- and field-tested but are not being actively marketed at present.

The economic evaluation of the use of additives for waste stabilization is site specific, at best, and must take into account not only the applicable disposal regulations, but also the type of waste and the disposal area hydrogeology. In some cases, for example (dry impoundments) it may be possible to stabilize materials to form containment dikes and basal layers into which unstabilized materials could be placed. This would, of course, depend upon the handling properties of the untreated wastes.

Table 5

Waste Treatment Processes

Process Type	Supplier/Developer	Additives	Degree of Commercialization	Commercial Waste Application	Level of Testing/Operation with FGD Wastes
Cement (Lime)-Based	IUCS, Inc.	Ash & Lime	Commercial/U.S.	FGD, Fly Ash	Full-Scale
	Dravo Lime Co.	Slag & Lime	Commercial/U.S.	FGD, Fly Ash, Mine Tailings	Full-Scale
	TJK, Inc.	?	Commercial/Japan	Industrial Inorg., Dredge Spoils	Full-Scale (Japan)
	Chem-Nuclear System, Inc.	Cement	Commercial/U.S.	Utility Radwastes	Unknown
	Commonwealth Edison/American Admixtures, Inc.	Lime & Fly Ash	Commercial/U.S.	FGD	Full-Scale
	Aerojet Liquid Rocket	Cement	Laboratory (FGD)	--	None
	Sludge Fixation Technology, Inc.	?	Laboratory (FGD)	Industrial Heavy Metal Sludges	None Reported
	Research Cottrell	Lime & Fly Ash	Field (FGD)	None	Field
	Envirotech (Chemfix)	Lime & Fly Ash	?	None	Lab?
Self-Cementing (plaster of paris)	Marston Associates	None	Pilot (FGD)	--	Field
Silicate-Based	Environmental Technology Corp.	?	Commercial/U.S.	Metal Hydroxide	None Reported
	Envirotech (Chemfix)	Silicate + Cement	Commercial/U.S.	Org. & Inorg. Industrial, Sewage	Field
	Ontario Liquid Waste Disposal, Ltd.	?	Commercial/Canada	Industrial	Field
	United Nuclear Industries	Silicate + Cement	?	(Radwastes)	None
	Stablex Corporation	Silicates	Commercial/U.K.	Industrial	None Reported
Thermoplastic	Werner & Pfleiderer Corp.	Asphalt	?	Radwastes	None Reported
	Southwest Research Institute	Epoxy Composites	Laboratory	--	None
		Sulfur	Laboratory	--	None
Organic Polymer	Chem-Nuclear System, Inc.	UF	?	?	None
	AMEFCO Co.	UF	?	?	None
	TRW Systems	Polybutadiene	Laboratory	--	None
	Protective Packaging (Teledyne)	?	?	Nuclear Wastes?	Unknown
Inorganic Precipitation	Industrial Resources, Inc.	Waste Acid + Iron	Field?	?	Lab
Unknown	Wehran Engineering Corp.	--	Laboratory?	?	None

Notes: 1. This is a generic listing for all wastes
2. The list is a partial listing

Source: (7,8)

3.0 DISPOSAL OF NONRECOVERY FGD WASTES

3.1 Disposal Options

A number of methods are potentially available for the disposal of FGD wastes either on land or in the ocean. Applicability of disposal options for FGD wastes can be broadly categorized on the basis of the nature of the wastes and the type of disposal.

Table 6 lists potential disposal options for the various types of wastes. In this table sulfur is included as a potential waste product; however, it is more likely that sulfur as a final product from recovery FGD systems will be produced for utilization. More importantly, recovery FGD processes are likely to require prescrubber systems to remove particulates, chlorides, and other flue gas constituents which might contaminate absorbent liquors. Prescrubber blowdown from these systems will result in wastes analogous to the wastes from nonrecovery FGD systems (although in smaller quantities). Hence, in the future if recovery processes are used, it will thus reduce, not eliminate, FGD wastes.

At present, all FGD wastes generated are disposed of on land. To provide a perspective on the current state of FGD waste disposal, Table 7 presents the breakdown on current disposal practices for operational FGD systems on utility boilers as of November 1978.

In addition to the above commercially operating units, a number of FGD systems and associated disposal systems are in operation for testing, development and/or data gathering purposes. A list of such current field testing programs on FGD wastes and associated data on the systems involved is presented in Table 8.

3.2 Regulatory Considerations

The disposal of FGD wastes is subject to regulations at both Federal and state levels. State regulations governing waste disposal on land can be more stringent than corresponding Federal regulations. At present, FGD wastes are disposed of exclusively on land. Ocean disposal may be a technically feasible alternative. In the future, ocean disposal may be carried out to a limited extent in regions where there are no mines available and disposal sites for land impoundments are scarce.

Disposal on Land

There are four major impact issues concerning land disposal:

- Waste stability/consolidation;
- Groundwater contamination;
- Surface water contamination; and
- Fugitive emissions.

Table 6

FGC Waste Types Versus Potential Disposal Options

Basis: All potential methods for disposal of wastes from particulate control and flue gas desulfurization methods are listed.

		LAND DISPOSAL						OCEAN DISPOSAL					
					Surface Mine Disposal			Shallow			Deep		
NO.	WASTES	Ponding With Water	Impoundment W/O Water Cover	Land Fill	Pit	Spoil Bank	Underground Mine	Conventional	Concentrated	Dispersed	Conventional	Concentrated	Dispersed
1.	UNSTABILIZED												
	Sulfite Rich and Mixed Sulfite/Sulfate												
	By Itself	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓
	With Ash	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓
	With Soil	NA	NA	✓	NA	NA	NA	NA	NA	NA	NA	NA	NA
	With Tailings	NA	NA	NA	✓	✓	✓	NA	NA	NA	NA	NA	NA
	Sulfate Rich												
	By Itself	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓
	With Ash	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓	✓
	With Soil	NA	NA	✓	NA	NA	NA	NA	NA	NA	NA	NA	NA
	With Tailings	NA	NA	NA	✓	✓	✓	NA	NA	NA	NA	NA	NA
	Sulfur	NA	NA	✓	✓	✓	✓	NA	NA	NA	NA	NA	NA
2.	STABILIZED												
	Sulfite-Rich and Mixed Sulfite/Sulfate												
	Soil Like	NA	✓	✓	✓	✓	NA	✓	✓	✓	✓	✓	✓
	Concrete Like	✓	✓	✓	NA	NA	✓	✓	✓	✓	✓	✓	✓
	Sulfate Rich & Gypsum												
	Soil Like	NA	✓	✓	✓	✓	NA	✓	✓	✓	✓	✓	✓
	Concrete Like	✓	✓	✓	NA	NA	✓	✓	✓	✓	✓	✓	✓

Notes: ✓ = Applicable
NA = Note Applicable

Source: Arthur D. Little, Inc.

Table 7

Summary of Disposal Practices for Operational FGC Systems on Utility Boilers as of November 1978

		Number of Plant/Plant Capacity								
		Western				Eastern				
			Wet Pond				Wet Pond			
Waste Form	System Type	Dry Fill	Lined	Unlined	Total	Dry Fill	Lined	Unlined	Total	TOTALS
FGD Waste Only	Lime-Based	--	--	--	0/0	--	--	--	0/0	0/0
	Limestone-Based	--	--	1/200	1/200	--	1/225	1/140	2/365	3/565
	Total	0/0	0/0	1/200	1/200	0/0	1/225	1/140	2/365	3/565
Codisposal	Lime-Based	2/865	--	2/960	4/1825	1/65	--	3/850	4/915	8/2740
	Limestone-Based	1/1585	2/2040	4/2460	7/6085	1/180	--	2/950	3/1130	10/7215
	Wet Particulate Scrubbing	5/1685	1/50	4/1195	10/2930	--	1/165	--	1/165	11/3095
	Total	8/4135	3/2090	10/4615	21/10840	2/245	1/165	5/1800	8/2210	29/13050
Stabilized FGD Waste	Lime-Based	--	--	--	0/0	3/1720	--	1/1650	4/3370	4/3370
	Limestone-Based	--	--	--	0/0	2/730	--	--	2/730	2/730
	Wet Particulate Scrubbing	--	--	--	0/0	1/165	--	--	1/165	1/165
	Total	0/0	0/0	0/0	0/0	6/2615	0/0	1/1650	7/4265	7/4265
TOTALS		8/4135	3/2090	11/4815	22/11040	8/2860	2/390	7/3590	17/6840	39/17880

Source: Arthur D. Little, Inc.

TABLE 8

SUMMARY OF CURRENT FIELD TESTING PROGRAMS FOR FGC WASTE DISPOSAL

Basis: Status as of November 1978

Location Utility (Plant)[a]	Sponsor(s)[a]	Principal Contractor(s)[a]	Scrubber System		Waste Characteristics		Disposal Mode	Test Area	Program Status
			Mode	Type	Type	Form			
LAND DISPOSAL:							Surface Mine	Section of Mine	Underway
Minnkota Power (M.R. Young)	EPA	UND/ADL	SO_2 Only	Alkaline Ash	Sulfate-Rich	Filter Cake (Unstabilized)	Stacking	~1-Acre Area	Underway
Gulf Power (Scholz)	EPRI	CIC/Radian	SO_2 Only	Limestone	Gypsum	Thickened Slurry (Unstabilized)	Dry Impoundment	1-Acre Pit	Planning
Gulf Power (Scholz)	EPA/EPRI	CEA/ADL	SO_2 Only	Dual Alkali	Sulfite-Rich	Filter Cake (Stabilized & Unstabilized)	Dry Impoundment	50-Acre Site	Planning
Columbus & S. Ohio (Conesville)	EPRI	MBA/Battelle	SO_2 Only	Lime (Thiosorbic)	Sulfite-Rich	Filter Cake (Stabilized)	Dry Impoundment	Small Pits/Ponds	Underway
Louisville Gas & Electric (Paddy's Run)	EPA	CE/LGE/UL	SO_2 Only	Lime (Carbide)	Sulfite-Rich	Filter Cake (Stabilized & Unstabilized)	Dry Impoundment	?	Planning
Louisville Gas & Electric (Cane Run)	EPA	Bechtel	SO_2 Only	Dual Alkali	Sulfite-Rich	Filter Cake (Stabilized)	Wet & Dry Impoundments	6 Pits (<.1 Acre)	Underway
TVA (Shawnee)	EPA/TVA	TVA/Bechtel	SO_2 & SO_2 + Ash	Lime & Limestone	Sulfite-Rich	Filter Cake Centrifuge Cake Thickened Slurry (Stabilized & Unstabilized)	Dry Impoundment	4 Pits/Area	Underway
			SO_2 Only	Limestone (Forced Oxidation)	Gypsum	Filter Cake			
OCEAN DISPOSAL:							Reef Construction	~0.2 Acre	Underway
Duquesne (Elrama/Phillips)	DOE/EPA/EPRI/NYSERDA/PASNY	SUNY/IUCS	SO_2 + Ash	Lime (Thiosorbic)	Sulfite-Rich	Filter Cake (Stabilized)	Concentrated Dump	1/2-Acre Pond	Underway
---	EPA	NEA/ADL	SO_2 & SO_2 + Ash	Many	Many	Many			

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[a] ADL - Arthur D. Little
CE - Combustion Engineering
CEA - Combustion Equipment Associates
CIC - Chiyoda International
DOE - Department of Energy
EPA - U.S. Environmental Protection Agency
EPRI - Electric Power Research Institute
IUCS - IU Conversion
LGE - Louisville Gas and Electric
MBA - Michael Baker Associates
NEA - New England Aquarium
NYSERDA - New York State Energy Research & Development Authority
PASNY - Power Authority of the State of New York
SUNY - State University of New York
TVA - Tennessee Valley Authority
UL - University of Louisville
UND - University of North Dakota

These are essentially regulated under the federal legislative framework listed in Table 9. All these legislative acts impose an element of constraint on FGD waste disposal. However, the Resource Conservation and Recovery Act is the major federal environmental legislation regulating disposal in mines, landfills and impoundments. According to proposed regulations under RCRA, Section 3001 defines criteria for determining whether wastes are hazardous or not. The most pertinent of these for FGD wastes are the toxicity-related tests. If a waste fails these tests, the disposal of that waste would be regulated under Section 3004 of RCRA. If an FGD waste fails these tests, its disposal would be regulated as a special case under Section 3004. If a waste passes these tests, the waste would be considered nonhazardous. Proposed regulations include guidelines under Section 4004 for nonhazardous waste disposal. Further definition of regulations including design standards and criteria are expected.

Disposal in the Ocean

Regulation of dispersed ocean dumping of stabilized and unstabilized FGD waste falls under the Marine Protection Research and Sanctuaries Act and is administered by the Environmental Protection Agency. If stabilized, brick-like FGD waste is used to create artificial fishing reefs with EPA concurrence, the activities would not be subject to ocean disposal criteria.

3.3 Land and Ocean Disposal Methods

Land Disposal

The principal methods of land disposal are:

- Wet ponding;
- Dry impoundment; and
- Mine disposal.

Wet Ponding: This method is at present more widely used than any other. Ponding can be employed for a wide variety of FGD wastes including unstabilized materials; however, ponding has been employed with the Dravo stabilization process. Ponds can be designed based on diking or incision and can even be engineered on slopes. But the construction of dikes or other means of containment for ponds is usually expensive. In the future, particularly if stabilization of FGD wastes is widely practiced, ponding will probably be limited to those sites that can be converted to a pond with minimal construction of dams or dikes. A special case of wet ponding is gypsum stacking now under evaluation. In this case, if the operation were analogous to that for phos-gypsum, gypsum slurry (typically from forced oxidation systems) would be piped to a pond and allowed to settle and the supernate recycled. Periodically the gypsum would be dredged and stacked around the embankments, thus building up the entrainment.

Table 9

Major Regulatory Framework For FGD Waste Disposal

Impact Issue	Legislation	Administrator
Groundwater Contamination	Resource Conservation and Recovery Act of 1976	Environmental Protection Agency, Office of Solid Waste
	Safe Drinking Water Act of 1974	Environmental Protection Agency, Office of Water Supply
Surface Water Contamination	Federal Water Pollution Control Act Amendments of 1972	Environmental Protection Agency, Office of Water Programs
	Marine Protection Research and Sanctuaries Act	Environmental Protection Agency, Office of Marine Protection
Physical Stability	Surface Mining Control and Reclamation Act of 1977	Office of Surface Mining Reclamation and Enforcement, Department of Interior
	Dam Safety Act of 1972	Army Corps of Engineers, Department of Defense
	Federal Coal Mine Health and Safety Act of 1969	Mining Enforcement Safety Administration, Bureau of Mines, Department of Interior
	Occupational Safety and Health Act of 1970	Occupational Safety Health Administration, Department of Labor
Fugitive Air Emissions	Clean Air Act of 1970 and its Amendments of 1977	Environmental Protection Agency, Office of Air Programs
	Federal Coal Mine Health and Safety Act of 1969	Mining Enforcement Safety Administration, Department of Labor
	Occupational Safety and Health Act of 1970	Occupational Safety and Health Administration, Department of Labor

Leaching from wet ponds is likely to be an important environmental issue that must be addressed in pond design and operation. Recent R&D efforts on wet ponding have centered on:

- Most effective means of containing pollutants within the disposal area; i.e., study of potential liner material.
- Better definition of leaching mechanism from lined and unlined ponds.

Among the more important studies on liner materials relating to the disposal of FGD wastes are:

a) The U.S. Army Corps of Engineers Waterways Experiment Station (WES) is conducting a program to: (1) determine the compatibility of 18 liner materials with flue gas cleaning (FGC) wastes and associated liquors and leachates; (2) estimate the length of life for the liners; and (3) assess the economics involved with purchase and placement (including disposal area construction) of various liner materials. The liners that WES is testing include:

    i) Admixture types (cement, lime, fly ash)
    ii) Prefabricated liner membranes (polymer, neoprene coated, etc.)
    iii) Spray-on types (polyvinyl acetate, latex, asphalt, cement, etc.)

    Results of this investigation are expected to be available in 1979.

b) In 1978 EPRI initiated a program to evaluate leachate control and monitoring systems for solid waste disposal facilities. The objective is to evaluate liner materials for utility solid wastes. This 36-month program which will be underway in 1979 is expected to yield substantial technical data on a number of liner materials.

Proposed RCRA Guidelines (9) provide some indication of potentially appropriate impoundment design. The extent of leaching of pollutants from disposal ponds is dependent on several factors: the hydrostatic head in the pond, which forces percolation through the pond bottom; the nature of the waste--primarily its permeability and the solubility of contaminants it contains; and the characteristics of the soil around the pond. At present, monitoring wells exist in some of the FGD waste disposal ponds, particularly the larger, more recent ones like the disposal pond for the Bruce Mansfield plant. However, insufficient time has passed for adequate data to be available. Efforts are continuing in this field and additional insights on field-scale leaching is likely to be available in the future.

. Dry Impoundment Methods: These may include any of the following variations:

- Interim ponding followed by dewatering and sometimes excavation and landfilling;
- Mechanical dewatering and landfilling of FGD wastes;
- Blending with fly ash and landfilling of FGD wastes; and
- Stabilization through the use of additives (non-proprietary or otherwise).

Typically, for dry impoundment type of disposal, the wastes are thickened and dewatered to a high solids content level and blended with fly ash and lime, thus forming a material with cementitious properties. This material is transported to the disposal site where it is spread on the ground in 1 to 3 foot lifts and compacted by wide track dozers, heavy rollers or other equipment. Layering proceeds in 1 to 3 foot lifts in segments of the site. The ultimate height of a disposal fill is site-specific but may be 30 feet to as high as 80 feet or more. A properly designed and operated dry impoundment system can potentially enhance the value of the disposal site after termination or at least permit post operational use.

Important engineering considerations in dry disposal of FGD wastes in landfills include:

a) Physical instability is a potential problem for all FGD wastes, including stabilized materials. Geometric factors such as height and slope angle in a landfill are interrelated; stability depends on the combination of fill height, slope angle, wastes density, degree of saturation, effective cohesion, effective angle of shearing resistance, and behavior during shearing (dilatant vs. densifying). For a given material, safe fill height decreases with increasing slope angle. For proper design, data is required in which maximum safe fill height is related to slope angle and soil shearing behavior. Such relations are not fully developed yet for FGD wastes because adequate data from proper tests (triaxial compression tests on consolidated samples with measurement of porewater pressures) have not been available.

   Instability problems may be ameliorated by compaction since compaction may produce several changes: voids may be eliminated (between chunks, not between individual particles); waste density may be increased; particles moved closer together may be bonded more effectively in stabilization reactions; and effective stress and residual total stress levels may be increased. If the wastes resemble sandy soils, an increase in density during compaction may create an important increase in angle of shearing resistance.

b) Climate could affect the stability of a wastes deposit in several ways. The total amount of rainfall and the intensity of rainfall both affect surface erosion, but these factors influence the rate of erosion; i.e., erodible materials will erode if wind and rain act on them but increase in amount/duration and intensity of wind and rain will increase the rate of surface erosion. Of course, as mentioned above, infiltration of water into wastes deposits can lead to saturation, decrease in apparent strength, and, possibly, to failure. This would be most likely to occur with wastes of high permeability in an unsaturated condition, but it could also occur in any FGD waste deposit.

Freeze-thaw cycles may also impact waste strength and stability. Stabilization could be retarded, disrupted or destroyed by an episode of freezing soon after mixing and placement. Freezing could produce cracks in the near-surface layers of a wastes deposit (frost polygon behavior). Cracks in the wastes deposit could yield greater mass permeability and infiltration rate even though waste blocks between cracks had been compressed and even dewatered.

c) Post Closure Land Use: Often it is difficult to specify the ultimate use of the land on which FGC waste disposal is planned or practiced. However, it is important to recognize engineering constraints on post-closure land use. Post-closure land use would tend to be limited by nature of the loads created or by the sensitive nature of the structures or facilities built on the wastes fill. For example, placement of some sort of fill in a uniform layer over the entire waste deposit should be feasible, but imposition of concentrated loads (e.g., footings in a building) may not be feasible.

d) Run-Off: To limit the impact of water run-off from waste deposits, grading and drainage are required as in any other type of construction activity. Consideration should also be given to run-off control following completion of the fill, including proper overall site grading and soil cover. Surface run-off caught in temporary retention basins could be recycled or discharged after appropriate treatment.

Mine Disposal: A disposal method that is receiving increased attention is mine disposal. It appears that surface coal mines and underground room and pillar mines for coal, limestone or lead/zinc ores offer particular potential (10). Of the four categories of mines noted above, coal mines, and in particular surface area coal mines, are the most likely candidates for waste disposal. Coal mines offer the greatest capacity for disposal, and they are frequently tied directly to power plants. In fact, many new coal-fired power plants are "mine-mouth" (located adjacent to the mine within a few miles) and the mine provides a dedicated coal supply. Since the quantity (volume) of FGC wastes produced is considerably less than the amount of coal burned, such mines usually would have the capacity for disposal throughout the life of the power plant.

In general, inactive surface mines are considerably less promising than active mines for FGD waste disposal. Unreclaimed surface mines can be used for disposal of wastes between remaining spoil banks, and these may offer suitable sites for disposal. However, because of recent surface mine reclamation legislation, the number of sites and total capacity for wastes available in the future will be limited.

In active surface mines, there are basically three options for the placement of FGD wastes:

- In the working pit, following coal extraction and prior to return of overburden;
- In the spoil banks, often return of overburden but prior to reclamation; and
- Mixed with or sandwiched between layers of overburden.

In any disposal operation in an active mine, though, the general overriding consideration is that disposal should cause minimal disruption of mining or reclamation activities. This provides a number of constraints on the disposal system.

a) First, the amount (volume) of sludge disposed of in any surface mine should not greatly exceed the amount of coal removed. The objectives of strip mine reclamation include returning the mined terrain to topographic configurations similar to original terrains. Also, returning significantly more waste to a mine than the coal extracted could slow down the mining and reclamation activities. In most cases, this does not really represent a constraint, since the wastes returned to the mine will usually be only those resulting from the coal removed. The amount of FGD waste generated from the combustion of coal will be considerably less than the quantity (weight or volume) of coal removed. Depending upon the type of coal, FGC system, and emission standards to be met, the volume of total waste (ash plus calcium-sulfur solids) will range from less than 10% of the coal burned to slightly over 50%. With Western coals, which are relatively low in sulfur, less than 20% is the rule. With higher sulfur Eastern and Interior coals, the total amount of wastes will typically run 30% or more.

b) Second, the physical condition (consistency) of the wastes must be amenable to ease of handling, transport, and placement using earth-moving equipment with minimal potential impact on the mining operations. For pit-bottom disposal this means that the wastes at the time of placement or immediately thereafter should have as a minimum the consistency of a soil-like material with little or no liquefaction potential. A slurry-like material or a waste with a tendency to flow either when placed or when overburden is dumped on top would present significant operational problems or unacceptably high costs for containment measures. A little more leeway exists for disposal in V-notches (between spoil banks); however, here again, soil-like materials or relatively cohesive materials that are relatively easily handled and transported will result in the least cost and minimal disruption of reclamation activities. At the least, the wastes must be well filtered (55% solids or higher), and may need to be admixed with dry fly ash, or stabilized.

c) Finally, minimal use should be made of existing mine equipment for transport and placement of the wastes at the mine. Dedicated equipment is preferred and, in most cases, mandatory. In almost all scenarios, waste is most easily placed by truck dumping. The use of coal trucks for this purpose could lead to unacceptable delays in coal mining operations due to the additional time for waste loading and discharging (and possible cleaning operations) Furthermore, most large mines use large bottom-dump trucks for coal haulage. These are designed to carry as much as 150 tons of coal and are usually constructed of aluminum. Not only can the waste corrode the aluminum, but the bottom dumping of wastes would be impractical. These trucks are not designed for ease of maneuvering, and operating them (or any other equipment) on a freshly dumped layer of waste would be difficult at best. The type of truck used for transport and placement will greatly affect the quantity of waste that can easily be disposed of.

At present there are only two commercial operations involving mine disposal of FGD wastes in surface coal mines--one at Texas Utilities' Martin Lake Station and the other at Square Butte's Milton R. Young Station (North Dakota). Both stations fire lignite and involve returning combined fly ash and calcium-sulfur solids from SO_2 removal to the respective mines. The operation at the Baukol-Noonan mine which supplies coal to the Milton R. Young Power Station is currently an EPA mine disposal demonstration project. At this mine both pit-bottom and spoil bank disposal are being employed. Mine disposal of FGD wastes can potentially be employed for subsidence control. Acid mine drainage neutralization, reclamation of mine areas or as soil amendments for tailings disposal from mining operation. Thus, there could be subsidiary benefits from this type of disposal.

Ocean Disposal

Ocean disposal of FGD wastes is not practiced today. However, if it could be practiced under environmentally acceptable conditions, it could represent an important option, particularly in Federal Regions 1 and 2 (the Northeast) where land for disposal is limited. For this and other reasons, EPA has been studying the disposal of FGD wastes in the ocean (10). From a technical viewpoint, ocean disposal may be considered in the shallow ocean (i.e., on the continental shelf) or deep ocean (off shelf). Each of these has a different ecosystem with a different set of potential impacts. At present a number of viable techniques exist for transporting FGC wastes to offshore disposal sites. These include:

- Self-propelled hopper ship with throttled discharged disposal or with a sudden dump capability;
- Tow-barge transportation and controlled dispersal over a great expanse of water or a sudden total bottom dump; and
- Submarine pipeline transportation and dispersal at a preselected offshore disposal site.

All of these approaches are technically feasible systems which have been utilized on a full scale for disposal of other types of wastes. Selection among them depends upon both the characteristics of the FGC wastes to be dumped and any environmental conditions and/or constraints that might exist.

At present, regulation of dispersed ocean dumping of stabilized and unstabilized FGD waste falls under the Marine Protection Research and Sanctuaries Act and is administered by the Environmental Protection Agency. The dumping would be required to be limited to an EPA-prescribed dumpsite under the following conditions:

- Trace contaminant (e.g., Hg, Cd) content of the dumped materials would be no higher than 50% above that of background sediments at the dumpsite;
- Concentrations of the dumped material in the water column four hours after release would not exceed 1% of the 96-hour LC_{50} of the material to local sensitive species; and
- No feasible alternatives to ocean disposal are available.

Stabilized, brick-like FGD waste may potentially be used to create artificial fishing reefs with EPA concurrence. Artificial fishing reefs are not subject to ocean disposal criteria but FGD waste disposal may be a special case. While ocean disposal of FGD sludges is an option that is perhaps available to throwaway system users with economic access to the ocean, new ocean disposal initiatives are now discouraged by the regulatory agencies.

At present, the major studies under EPA sponsorship or participation relating to the ocean disposal of FGD wastes are:

a) The Arthur D. Little study (10) for EPA on the technical, economic and environmental feasibility of the ocean disposal of stabilized and unstabilized FGD wastes. This study includes both laboratory and small-scale field testing relating to impact issues.

b) The State University of New York (11) with funding provided from Power Authority of the State of New York, New York ERDA, DOE, EPRI, and EPA, is studying the use of stabilized brick-like FGD wastes (using IU Conversion Systems process) to create artificial reefs for marine habitats. This study is expected to continue for 2 to 5 years.

3.4 Disposal Options Versus Potential Impact Issues

On balance, technology exists for environmentally sound disposal of FGD wastes. The environmental impact issues requiring consideration in handling and disposal of FGD wastes are:

- Air-related. These include fugitive particulate emissions, emissions of SO_2 and H_2S and emissions of trace metal compounds;
- Water-related. These include groundwater contamination, surface water point source discharges and run-off;
- Land-related. These include physical stability (subsidence, liquefaction or other structural failure, erosion, etc.) and land use considerations; and
- Biological impacts both in the site and adjacent areas and consequential effects.

Potential impact issues are highly site- and system-specific. With that understanding, Table 10 illustrates the major types of impact issues associated with various disposal options. As the matrix illustrates, the range of waste types and possible disposal conditions is sufficiently broad to eliminate the potential for "generally significant" issues to be associated with any of the disposal options. Further, site-specific application of appropriate control technology can be employed to mitigate adverse impacts. In other words, issues of potential significance in FGD waste disposal can best be defined in terms of specific waste types, disposal practices, and disposal environments. The significance of many potential impact issues may be better quantified by additional field-scale operating experience (and environmental monitoring) with FGD waste disposal. As indicated in the matrix, this is particularly desirable for defining potential issues in the categories of water quality and biological impacts.

3.5 Assessment of Present Control Technology

It is expected that much of the difference between potential and actual impacts for the FGD waste disposal options discussed above will be determined by the degree to which presently available control technology becomes incorporated as "good design" and "good practice" in typical disposal operations. Good design and practice could also minimize the potential for adverse impact from abnormal events. Important considerations in the application of present control technology are briefly discussed below.

a) Site Selection: Site selection may or may not be considered control technology. However, there is no question that proper site selection could by itself ameliorate or eliminate most of the potential disposal impacts discussed above. Specifically, the following mitigative combinations of site characteristics and impact issue categories are considered applicable:

Table 10

Disposal Options Vs. Potential Environmental Impact Issues for FGC Wastes

	Potential Environmental Impact Issues				
Disposal Options	Land Use	Surface Water Quality	Groundwater Quality	Air Quality	Biological Impact
Wet Ponding	2	2*	2*	3	2*
Dry Disposal	2	2*	2*	2*	2*
Surface Mine Disposal	2	2*	2*	2*	2*
Underground Mine Disposal	3	3	2*	3	2*
Shallow Ocean Dumping	3	2*	3	3	2*
Deep Ocean Dumping	3	2*	3	3	2*

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* Significance highly uncertain due to data gaps.

Key: 1 = Issue of potential general significance for all FGC wastes at all disposal sites.

2 = Issue of potential significance for specific types of FGC wastes and/or specific disposal sites.

3 = Issue of minor or no potential significance.

Potential Impact Issue	Mitigative Site Characteristics
Land Use	Proper topography, geology and hydrology; absence of nearby conflicting land uses.
Water Quality	As above for land use, plus absence of nearby sensitive receiving waters (surface or aquifers). For example, a small stream or very pure aquifer may impose greater constraints than a relatively large stream or impure aquifer.
Air Quality	Absence of "non-attainment area" and Class I Prevention of Significant Deterioration designations for total suspended particulates. Usually this is even more important for the Power Plant Siting.
Biological Effects	Absence of sensitive biological resources.

b) Dewatering: As discussed earlier, dewatering of FGD waste prior to processing or land disposal can result in major improvements in physical stability and reduce water quality impacts regardless of which disposal approach is employed, including those discussed below.

c) Stabilization: Stabilization appears to be highly relevent to the mitigation of land use issues, including the potential for abnormal events (i.e., disposal area liquefaction or other catastrophic failure modes), and the suitability of disposal sites for a broader range of post closure uses requiring increased bearing strength. Stabilization techniques resulting in decreased waste permeability can be considered mitigative of potential water quality impacts due to leachate migration. This factor should be considered in balance with the requirements for disposal area runoff control on a site-specific basis.

Stabilization reduces permeability and hence reduces rate of contaminant transfer from a disposal site. However, long-term cumulative contaminant migration could be important. In particular, it is not clear that reductions in long-term trace contaminant availability would take place when fly ash is used as a stabilization additive to a waste initially containing no ash. However, migration of contaminants to the environment at a slower rate is more desirable.

Cementitious stabilization processes, because of increased particle size, may also be considered mitigative of the potential for post-disposal fugitive particulate emissions from dry FGC waste disposal operations, and may minimize or prevent gaseous emissions by reducing exposure of waste to water and biological organisms.

In ocean disposal, cementitious stabilization may remove liabilities of FGC wastes as benthic substrates and as sources of sulfite-related depletion of dissolved oxygen. However, questions of sulfite and trace contaminant availability, among others, preclude definitive judgment on this issue at this time.

d) Forced Oxidation: The intentional production of sulfate-rich, rather than sulfite-rich FGC wastes, is presently a subject of considerable interest. In ocean disposal, the sulfate-rich products of forced oxidation would have the obvious advantage of mitigating the potential for sulfite-related depletion of dissolved oxygen. This advantage would be shared in land disposal operations (especially wet impoundments), but its relative importance is less clear. A dominant question concerning the mitigative potential of forced oxidation for land disposal is whether or not the process results in increased or decreased physical stability. Based on experience with soils, gypsum FGD wastes comprised of relatively uniform, sand-sized particles may exhibit considerable failure potential in the absence of: 1) effective compaction and dewatering, and/or 2) co-disposal with materials of varying particle size (i.e., fly ash). However, if FGD gypsum is analogous to phos-gypsum, recrystallization mechanisms occurring in the disposal pile may improve stability.

e) Co-disposal of Wastes and Creation of Waste/Soil Mixtures: Although the term co-disposal is often used in reference to the creation of disposal mixtures of two waste streams (e.g., FGD wastes and coal ash), it is used here to imply a broader range of potential opportunities. Specifically, for land disposal of FGD wastes, "co-disposal" might also include the application of technologies for the creation of soil/waste mixtures. If soils with the proper characteristics are available, the creation of soil/waste mixtures may be an alternative to the addition of fly ash where only limited increases in physical stability are desired in a disposal operation, or where trace contaminant availability needs to be reduced to facilitate revegetation or decrease water quality impacts. Traditional co-disposal involving fly ash plus FGD waste appears to have substantial advantages over independent disposal in terms of improved physical stability and (potentially) decreased permeability. This might be especially relevant to sulfate-rich FGC wastes of uniform particle size. However, in some situations the extent to which the ash serves as a reservoir of certain trace contaminants could prove a liability from the standpoint of potential water quality degradation.

3.6 Future Research Needs

A number of programs have been undertaken (and are in progress) by the Environmental Protection Agency (EPA), the Department of Energy (DOE), the Electric Power Research Institute (EPRI), and others. These efforts have provided much of the baseline information for environmental assessment. Provided these programs continue, additional data and insight permitting better environmental assessment will be possible.

Research needs pertinent to environmental assessment of FGC disposal are:

a) Acquisition of field data on the actual impacts of full-scale disposal operations under varying environmental conditions. Field-scale monitoring of large disposal operations over a period of several years is warranted. EPRI's proposed program at Conesville Plant of Columbus and Southern Ohio Power is one such example. EPA is also planning an extensive 2-year study on characterization and environmental monitoring of full-scale utility disposal sites.

b) A corrollary of (a) above would be the development of correlations and tools of extrapolation to relate existing lab/pilot scale data on physical stability and water quality impacts to full-scale field data.

c) Integrated study and evaluation of the environmental trade-offs in co-disposal of various FGD wastes and various coal combustion ashes. (It appears that this type of initiative could emphasize laboratory work with limited pilot and full-scale field verification.)

d) Development of basic data (laboratory and field-scale) on the biological impact potential of principal land-based FGC waste disposal options, especially data relating to water-related impacts of major soluble species and trace contaminants. Typical questions are:

- What are the biological and health effects of mixtures of trace metals (in the form found in liquors), such as zinc, copper, lead, mercury, cadmium or nickel in combination with selenium in particular, but also in other combinations?
- What is the uptake of potentially toxic materials by vegetation and wildlife associated with disposal areas?
- What are the levels of ambient concentration of waste-related potentially toxic materials in vegetation and surface water that may produce chronic health problems for wildlife?

EPA is presently supporting biological testing work on FGC wastes at Oak Ridge National Laboratory.

e) Development of basic (laboratory and field) data on the potential for fugitive particulate emissions from areas previously used for the dry disposal of FGC wastes.

f) Socio-economic impacts of FGC waste disposal on land need to be better defined.

In the future, FGD waste generation will not be limited to those by utility systems. Coal utilization in industrial boilers (25 MW or larger) is also likely to grow substantially. FGD wastes from such industrial boilers (while analogous in composition to those from utility boilers) present additional waste management issues due to differences in distribution of generation facilities, in quantity of FGD wastes generated at each facility and other factors. These issues also require further evaluations and study.

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MARKETING ALTERNATIVES FOR FGD BYPRODUCTS

AN UPDATE

By

W. E. O'Brien and W. L. Anders
Emission Control Development Projects
Tennessee Valley Authority
Muscle Shoals, Alabama

ABSTRACT

The latest FGD byproduct marketing programs conducted by TVA for EPA are "First Annual Update - Potential Abatement Production and Marketing of Byproduct Sulfur and Sulfuric Acid in the U.S." and "Flue Gas Desulfurization Byproduct Production and Marketing System Users Manual."

The first is an annual update to 1983 of an extensive computer system to project FGD byproduct sulfur and sulfuric acid market potential. Results of this update are presented. Boilers projected to come on-line between 1978 and the end of 1983 are important candidates for marketing FGD byproduct sulfuric acid. A small reduction in total FGD byproduct sulfur costs could result in emergence of some competitive FGD sulfur production.

The users manual describes programing, procedures, and techniques required to utilize the computer production and marketing system. The users manual makes the byproduct production and marketing computer system available on a commercial basis through TVA under a technology transfer agreement with EPA. Procedures for access are described.

MARKETING ALTERNATIVES FOR FGD BYPRODUCTS

AN UPDATE

INTRODUCTION

For several years the Tennessee Valley Authority (TVA), in conjunction with the U.S. Environmental Protection Agency (EPA) and others, has conducted economic studies related to flue gas desulfurization (FGD) processes. As a part of these studies a computer-based byproduct marketing system was developed to evaluate the potential of marketing sulfur byproducts from FGD processes. It compares the economics of using clean fuel with the economics of several FGD processes, including limestone scrubbing with waste sludge production, magnesia scrubbing with sulfuric acid production, and the Wellman-Lord - Allied Chemical scrubbing process which produces sulfur. The system contains data on over 900 U.S. power plants and the locally applicable regulatory information required to calculate scrubbing costs for each power plant in the system. In the case of sulfuric acid production, the marketability of the acid is determined using data on U.S. sulfur and sulfuric acid transportation and the U.S. sulfuric acid manufacturing industry. Provisions for sulfur marketing are being incorporated in the computer system. General or specific cost determinations can be made for different alternative clean fuel levels (ACFL) and different scrubbing methods, including the possibilities and effects of byproduct marketing.

A scrubbing system that produces a potentially marketable byproduct such as sulfuric acid is not practical for all power plants because the incremental cost of recovery can exceed the market value. A scrubbing system that produces waste byproducts is not satisfactory in every case because of disposal problems, delivered raw material costs, and plant operating characteristics. Neither will the exclusive use of a clean fuel solve the problem for all plants. Consideration must, therefore, be given to engineering, economics, transportation, and marketing factors which allow in-depth analyses in selecting a suitable strategy for particular power plants. Since several strategies may be selected that result in compliance with clean air requirements, information must be available to make decisions that result in the least cost of compliance in the long run.

A definitive TVA report on a marketing analysis of byproduct sulfuric acid economics projected to 1978 was issued by EPA in 1978 (Bucy et al. 1978). Power plant abatement compliance decisions must, however, be made to conform with anticipated conditions several years in advance. The need for regulatory agency approval and longer construction periods combine to extend the effects of decisions into a more distant, and thus less certain, future. It is essential, therefore, that a project of this nature use the latest available data projected as far in the future as can be reasonably substantiated. For this reason an update of the 1978 projection for byproduct sulfuric acid market potential was recently projected through 1983 using data available up to September 1978 (O'Brien and Anders, in press). This update is a continuing refinement of the computer programs of the original byproduct marketing model. In addition, a manual analysis of byproduct sulfur marketing as an alternate to byproduct sulfuric acid is included.

A users manual (Anders, in press) was also prepared to make the byproduct marketing computer system available to the power industry on a commercial basis. This manual provides the information and procedures necessary to use the system. It is primarily intended for the analyst-programer but it should also be useful to those requiring detailed information about the system who do not have extensive systems or programing experience.

The byproduct marketing system is available through TVA under a technology transfer agreement with EPA. Procedures for releasing the system are initiated upon receipt of a written request. At the present time, under the same technology transfer agreement, selected runs of the system based on user-supplied data can be made by TVA. All inquiries concerning the byproduct marketing system and this manual should be directed to Emission Control Development Projects, Tennessee Valley Authority, Muscle Shoals, Alabama 35660, telephone No. (205) 383-4631, extension 2516.

SYSTEM DESCRIPTION

The users manual provides the only description of the complete byproduct marketing system. The system description presented herein is a condensed version of the comprehensive description presented in the users manual. The byproduct marketing system consists of a number of integrated computer programs, models, and data bases which can be used to make cost comparisons of FGD strategies designed to meet clean air regulations. For strategies which produce a salable byproduct, the marketability of the byproduct is determined and its effect on FGD costs is included in the cost comparisons. The system can use this data or user-supplied data to develop situations for comparison of alternate FGD strategies. For comparisons based on the use of clean fuel without FGD, an ACFL is used to reflect the cost differential between a complying fuel and a noncomplying fuel. In the cases of magnesia scrubbing and the Wellman-Lord - Allied Chemical process, the system determines the incremental cost of sulfuric acid or sulfur production. Incremental production cost is defined as the production cost per ton of sulfuric acid or sulfur above the cost of either limestone scrubbing or the ACFL value, whichever is lower. For sulfuric acid the system determines the marketability in terms of net back from sales to specific locations at which the delivery and incremental production costs are offset. Provisions for distribution of sulfur are being incorporated into the system.

The entire system can be used to compare the magnesia-scrubbing strategy with either the limestone-scrubbing or clean fuel strategies for any combination of geographic and power plant situations. The numerous subsystems, programs, and data bases can be used separately or in various combinations to provide information on a wide range of related FGD processes and on sulfuric acid manufacture, transportation, and marketing.

The byproduct marketing system is shown diagramatically in Figure 1. It can be divided into four subsystems. The supply subsystem consists of data bases and programs which provide data on power plants, emission control regulations, raw material costs, and FGD design and cost data. These can be used to determine scrubbing costs on a boiler-by-boiler basis for each power plant in the data base. The demand subsystem consists of programs and data bases on sulfur transportation costs and acid plant operating costs which are used to determine acid plant avoidable production costs. Avoidable production cost is the expenditure that could be avoided by shutting a sulfur-burning acid plant down. This savings is the break-even price that can be paid for FGD byproduct acid. The transportation subsystem consists of data bases and programs to provide rail mileages, tariffs, and rate-basing information for power plants and acid plants from the other subsystems. It is used to calculate acid transportation costs. The fourth subsystem consists of a linear program model generator, a linear program model solution generator, and various optional report generators. It uses the results of the other three subsystems to select the least-cost option for each power plant considered for marketing.

The data used in the system have been compiled from a wide range of U.S. Government, TVA, and published sources. The data include over 3500 boilers representing over 900 power plants and all acid plants and smelters in the United States. The automated model is limited to the eastern 37 states for which a railroad rate-basing system, necessary to the transportation subsystem, exists. Excess acid supply from the 11 western states and Canada is included in the linear programing model as a manually calculated factor. The calculation of scrubbing costs is based on design and economic premises developed by TVA and EPA to compare the economics of scrubbing systems (McGlamery et al. 1975).

The byproduct marketing system is designed on the basis of several assumptions important to a conceptualization of the evaluation process. They are listed as follows:

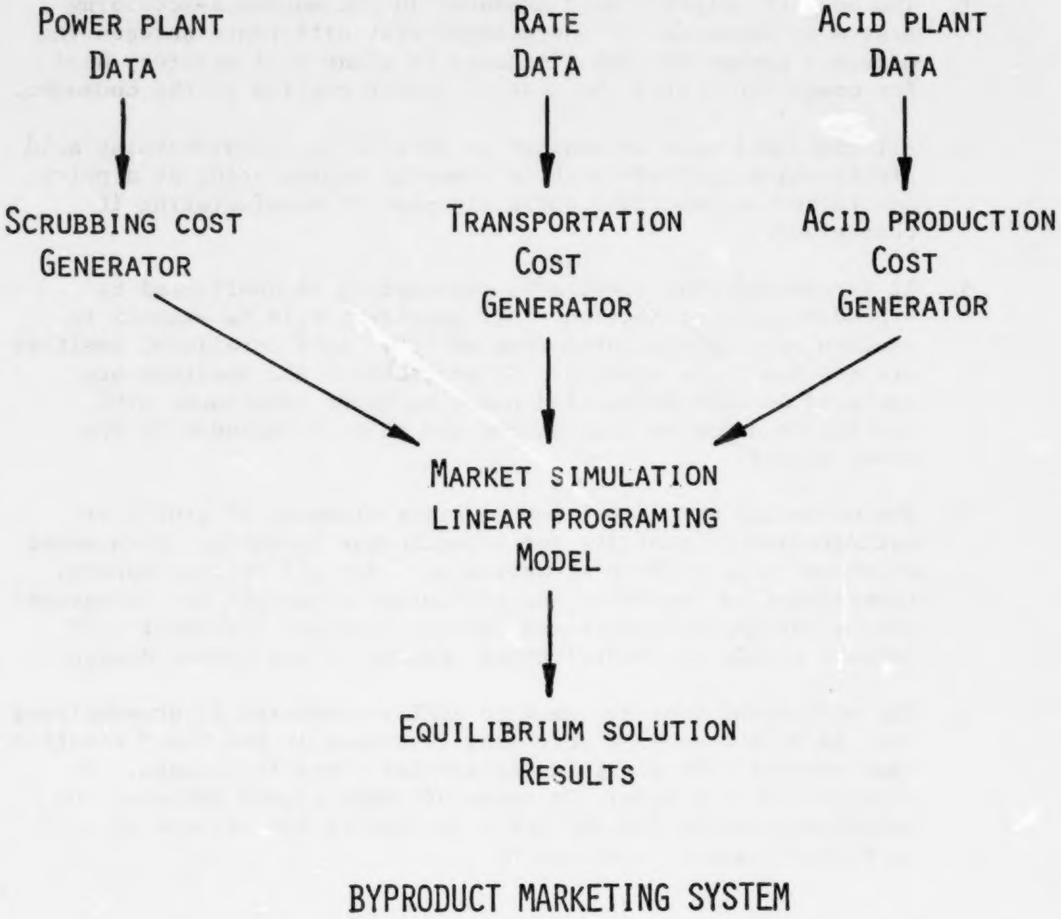
POWER PLANT DATA
RATE DATA
ACID PLANT DATA
SCRUBBING COST GENERATOR
TRANSPORTATION COST GENERATOR
ACID PRODUCTION COST GENERATOR
MARKET SIMULATION LINEAR PROGRAMING MODEL
EQUILIBRIUM SOLUTION RESULTS
BYPRODUCT MARKETING SYSTEM

Figure 1. Byproduct marketing system.

1. The cost of sulfuric acid produced in the magnesia-scrubbing system is based on the incremental cost difference between the magnesia system and the limestone or clean fuel strategy used for comparison, plus the cost of transportation to the consumer.

2. All byproduct acid is assumed to be sold to sulfur-burning acid plants which can reduce their costs by buying acid, at a price determined as described above, instead of manufacturing it themselves.

3. It is assumed that total acid consumption is unaffected by byproduct acid production. All byproduct acid is assumed to replace acid manufactured from sulfur. As a corollary, smelters are assumed to be producers of necessity. All smelters are assigned an acid production based on their compliance with applicable emission regulations and this is included in the total supply.

4. The marketing model does not consider elements of profit or maximization of benefits for a particular industry. Each model solution is an optimum situation in which all acid producers, transportation networks, and acid plant consumers are integrated into a system which provides for the greatest byproduct acid revenue within the restrictions imposed by the system design.

5. The additional cost assigned to ACFL as compared to noncomplying fuel is also used as a screening technique in the model construction process. By assigning particular clean fuel costs, the structure of the model, in terms of power plants included and comparisons made, can be varied to compare the effects of different costs of compliance.

FGD PROCESS DESCRIPTIONS

All of the FGD systems used in this evaluation are scrubbing processes in which the flue gas is contacted with a suspension or solution of absorbent in water. The SO_x in the flue gas reacts with the absorbent to form sulfur salts. A purge stream is removed and fresh absorbent added to maintain equilibrium concentrations in the scrubber system. All of the processes are in use or have been tested in full-scale operation (Herlihy 1977). The FGD systems are assumed to be installed downstream from existing air heaters and particulate matter removal equipment. All FGD equipment is provided, including raw material handling systems, auxiliary processing equipment to produce and store byproducts, and waste disposal facilities.

The limestone FGD system consists of a scrubbing system in which a suspension of finely ground limestone is contacted with the flue gas to form calcium sulfite and calcium sulfate. The purge stream is pumped to a disposal pond without further treatment. The system is characterized by large raw material and land requirements and relatively low energy, capital, and operating costs.

The magnesia FGD system is a similar scrubbing process using magnesia as the absorbent. The magnesium sulfite formed is removed from the purge stream, dried, and calcined to regenerate magnesia and sulfur dioxide. The magnesia is returned to the scrubber system and the sulfur dioxide is processed to sulfuric acid in an onsite acid plant. The system is characterized by lower raw material and land requirements and higher energy, capital, and operating costs, relative to the limestone system.

The Wellman-Lord - Allied Chemical process uses a sodium sulfite-bisulfite solution as the absorbent. Sodium sulfate is removed from the purge stream by selective crystallization. The bisulfite-rich purge is then thermally treated to regenerate sodium sulfite and sulfur dioxide. The sulfur dioxide is converted to sulfur by an onsite Allied Chemical proprietary reduction process. The characteristics of the system are similar to the magnesia system but raw material, energy, capital, and operating costs are higher for most situations.

FUNCTION AND USE OF THE USERS MANUAL

As the EPA-TVA byproduct marketing studies progressed, the automated system of programs and data bases that had been developed were expanded and refined in order to meet increasingly detailed requirements. It became apparent that, in addition to the direct benefits to the actual studies, the expanded system presented opportunities for others to utilize the resources that were being developed. An additional objective was defined in conjunction with the continuing studies to prepare a manual to describe how the system is used, and to make the complete system available for access on a commercial, nationwide, interactive time-sharing/remote batch data processing network when justified by potential usage.

The manual is designed for the needs of a programer-analyst; it is a how-to guidebook providing information and procedures necessary to use the system. It should also be useful to those requiring information about the system who do not have extensive systems or programing experience. It is primarily a users manual, however, and does not provide all of the concepts and background information necessary for the use of the system.

MODEL UPDATE

Following is a brief description of the major changes from the 1978 projection which were incorporated in the 1983 projection.

Scrubber Cost Generator

Power Plant and Boiler Screens. Past observations have shown that older, smaller power plants with small boilers are very poor candidates for FGD. In order to eliminate them from the model a screening method was adopted. It provided that individual boilers of less than 25-MW capacity or over 20 years old would automatically be eliminated from consideration. It also eliminated power plants of less than 100-MW total capacity. In this manner the model is run on a more cost-efficient basis and emphasis can be placed on the more logical candidates.

New Power Plants and Boilers through 1983. Projected boilers scheduled for completion between 1978 and the end of 1983 were added to the data base. There were 147 of these boilers identified and 26 were directly included in the 1983 model. The 121 projected boilers that were not considered are summarized as follows:

	Number of boilers not considered	Reason
	13	Location could not be identified.
	25	These boilers were located in the 11 western states where transportation rates were impractical to automate. They are included in summary form by manual analysis.
	23	An alternate scrubbing strategy was already selected.
	10	Fuel data were incomplete.
	49	A clean fuel strategy was already selected.
	1	The total plant size was less than 100 MW.
Total	121	

Updated Regulations and Compliance Plans. The 1978 projection was based on regulation data available from EPA through June 30, 1976. The 1983 projection used equivalent data through July 15, 1977, the latest available from EPA. It does not provide for regulation changes which may result from the revisions proposed in 1978 by EPA. The compliance status of power plants was also updated. A special report from EPA based on regulations assumed to be in effect December 31, 1978, was used as a basis.

Cost Escalation. Capital costs were escalated to 1983 using the *Chemical Engineering* cost indexes. The capital costs used in the 1978 projection were based on the same indexes. Operating costs were escalated to 1983 based on TVA projections.

The escalated cost data increased magnesia-scrubbing costs to a greater degree than limestone-scrubbing costs. The result was a relatively higher cost for magnesia scrubbing compared to limestone scrubbing in the 1983 projection.

Transportation Cost Generator

Rail Rate Increase. Rail rates were projected to 1983 based on the 1973-1977 5-year historical pattern and the anticipated effects of the "Railroad Revitalization and Regulatory Reform Act of 1976" (see Figure 2).

Barge Rate Increase. Barge rates were projected to 1983 based on the 1973-1977 5-year historical pattern, and the estimated effects of the passage of HR-8309. HR-8309 was a bill in the House of Representatives which had already passed the Senate at the time of the update report. This bill was signed by the President October 24, 1978, in essentially the same form used in this report. This legislation assesses a users tax on fuel used for commercial traffic on the inland waterways system as shown in Figure 3.

Transportation Cost Inflation. Transportation costs were projected to increase at an average annual rate of 13.4% from 1978 through 1983. This is an overall transportation cost increase of approximately 88% over the 1978 projection. It is significantly above other cost escalation rates and is expected to make transportation an increasingly important element of FGD byproduct marketing.

Acid Production Cost Generator

Sulfuric Acid Plant Demand Base. New sulfuric acid plant capacity between 1978 and 1983 was added to the demand data base. Plants projected to be shut down by 1983 were eliminated. Demand was equated to 75% of sulfur-burning sulfuric acid plant capacity. This update reflects actual fiscal year 1977-1978 production levels projected to 1983.

Sulfuric Acid Avoidable Production Cost. This cost, the expenditures in $/ton of sulfuric acid that could be avoided by shutting down the sulfur-burning acid plant, was modified substantially. The major change was an update of the byproduct steam credit to a sulfur-burning sulfuric acid plant. In addition, the price of sulfur at Port Sulphur was increased from $60/long ton to $70/long ton.

Market Simulation Linear Programing Model. The 1983 model was basically unchanged from the 1978 model with two exceptions: (1) The $0.35/MBtu (M = million) ACFL model was eliminated because of escalated

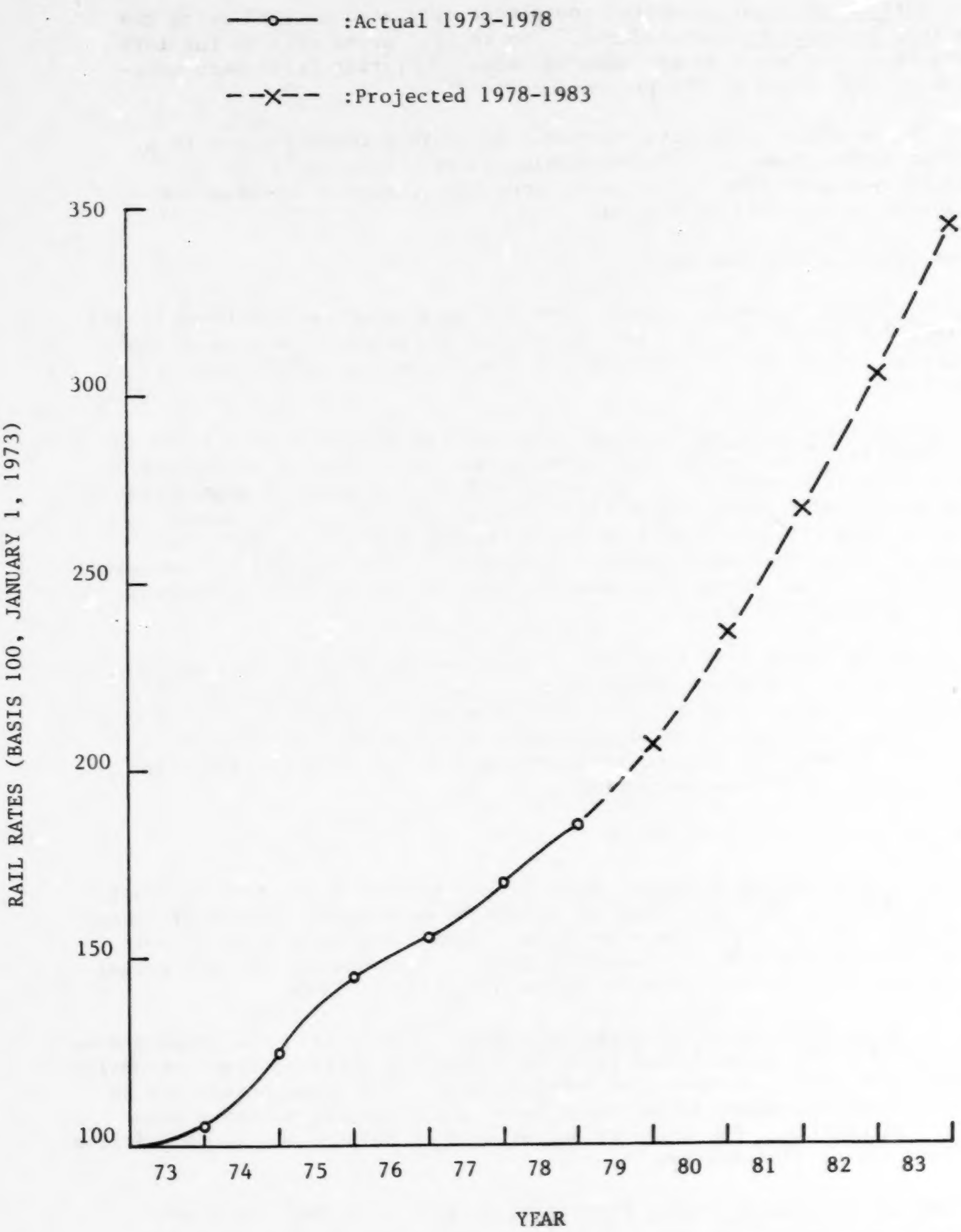

Figure 2. Rail rate increases, actual and projected, 1973-1983.

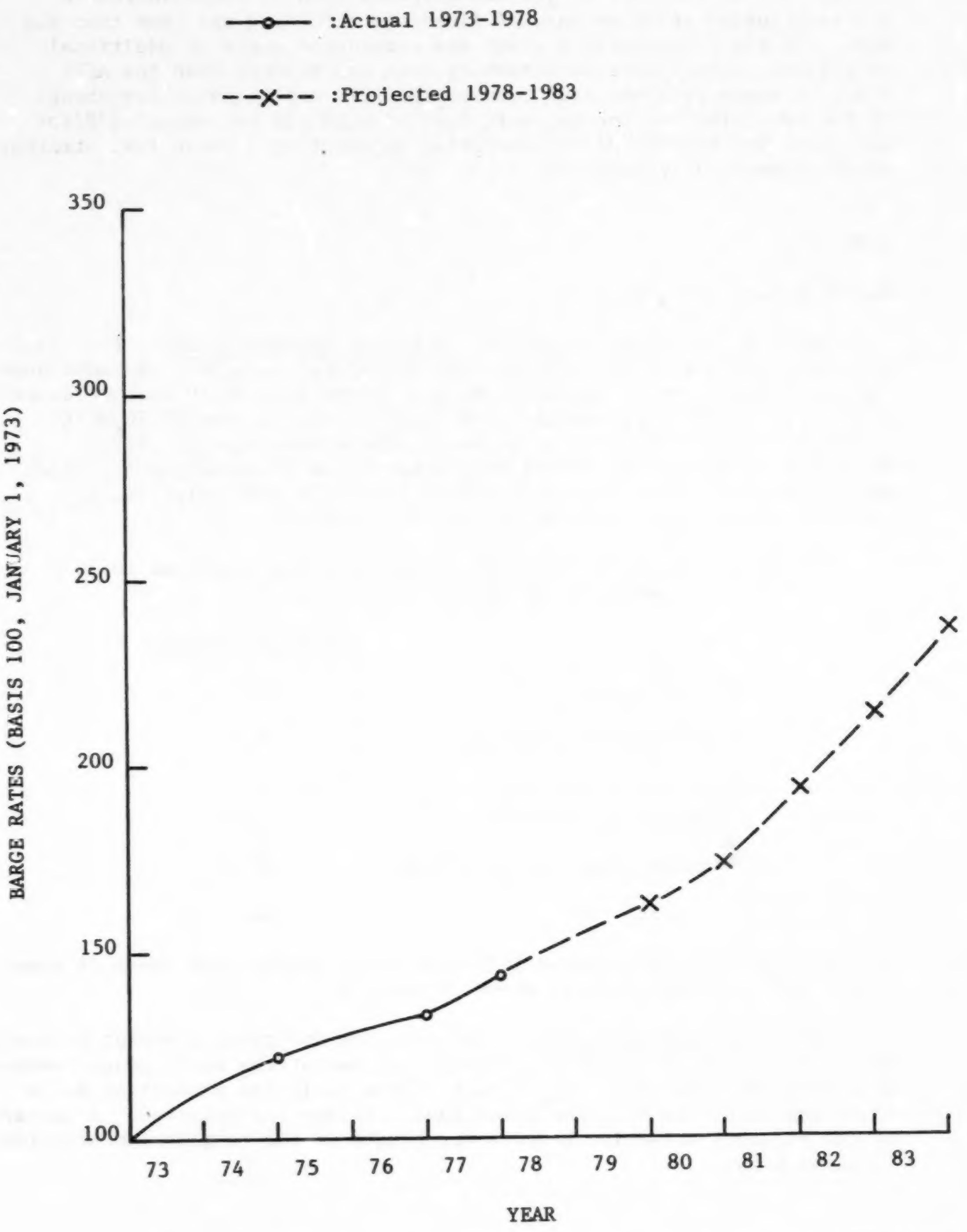

Figure 3. Barge rate increases, actual and projected, 1973-1983.

costs. (2) In the 1978 projection models a plant was considered in the equilibrium solution only if limestone scrubbing was less than the ACFL. In the 1983 models a plant was considered under an additional condition: when limestone scrubbing cost was greater than the ACFL a direct comparison was made between the ACFL and magnesia scrubbing. If the resulting incremental unit cost of acid did not exceed $30/ton the plant was considered for marketing rather than a clean fuel strategy being automatically selected.

RESULTS

FGD Byproduct Sulfuric Acid

Selection of Power Plants for Marketing Consideration. For 1983, 94 power plants with 165 boilers were calculated to be out of compliance and were considered in all model runs. In the 1983 $0.50 ACFL solution five power plants were considered for marketing. In the $0.70 ACFL solution 26 out of a total of 94 power plants were considered for marketing. Because the number of plants at the $0.50 ACFL was limited and these plants were also included in the $0.70 ACFL solution, a detailed analysis was limited to the $0.70 solution.

The $0.70 ACFL model generation prescreen that resulted in 26 plants being considered for marketing is shown below:

	Number of plants
All clean fuel	43
All limestone scrubbing	24
Mixed clean fuel and limestone scrubbing	1
Potential magnesia scrubbing	26
Total	94

The $0.70 ACFL solution potential power plant supply from these 26 power plants was 4,600,000 tons as shown in Table 1.

1983 $0.70 ACFL Solution. The solution indicated a market potential for 7 out of the 26 plants considered. At two of the seven plants market potential was limited to only a part of the projected production and a mixed magnesia-scrubbing and clean fuel strategy was selected. A summary of the strategy selection process for plants in the equilibrium solution is shown below:

Table 1. TWENTY-SIX POWER PLANTS INCLUDED IN 1983

$0.70 ACFL SOLUTION FOR ANALYSIS OF MARKET POTENTIAL

Plant No.	Location	MW considered	Incremental cost, $/ton acid	Tons of acid considered
0720000900	NC	1440	0.00	67,592
0785000500	IL	550	22.61	74,394
1040000200	OH	680	37.76	137,133
1095000200	OH	787	28.34	363,075
1095000600	OH	413	27.70	67,243
1115001300	IL	1271	26.31	321,051
1145001200	MI	1283	22.59	197,390
1145001900	MI	1185	13.09	117,946
1400000600	PA	525	51.66	69,209
1415000150	KY	800	26.27	132,361
1655000300	FL	1280	12.15	196,573
2185000900	TX	2820	0.00	156,384
2225000800	IL	2342	20.24	402,939
2260000100	IN	438	46.35	73,221
2260000500	IN	1239	25.96	156,607
2455000250	KY	2011	6.43	338,976
2755000600	KY	682	40.22	161,584
2755000650	KY	495	24.37	99,482
3080000400	MS	877	37.84	67,862
3795000350	PA	650	24.65	82,161
3840000500	PA	615	37.44	68,664
4045000250	IN	2587	13.39	334,927
4480000075	SC	595	29.51	100,205
4530000850	TX	951	27.77	159,425
4740000300	FL	1136	27.83	226,708
4820001800	MI	3247	13.26	449,826
				4,622,938

	Number of plants
Magnesia scrubbing	5
Mixed magnesia scrubbing and clean fuel	2
All clean fuel	1
All limestone scrubbing	18
Total	26

The seven power plants marketing sulfuric acid in the 1983 $0.70 ACFL solution had total sales of 1,243,000 tons as shown in Table 2. The ranked order shown is the result of application of four criteria: (1) the balance between the production capacity of the power plant and its potential market, (2) the balance between the acid capacity of each consumer and the byproduct acid supply from both power plants and smelters, (3) the potential sales margin per ton of acid from each power plant, and (4) the sales indicated as also occurring in the $0.50 solution.

Projected new boilers were a factor at six of the seven plants selected for marketing. They contributed almost 60% of the sales even though they made up only 15% of the total number of boilers considered for marketing.

The eastern smelter and Canadian supply was absorbed by the market in the 1983 equilibrium solution. No sales potential for the additional western supply through transshipment terminals was indicated because of higher transportation costs.

The solution would be fairly stable in the event of a sulfur price reduction. Less than 200,000 tons out of a total of 1,200,000 tons would be affected by a $20/long ton sulfur price reduction.

FGD Byproduct Sulfur

There is a considerable interest in elemental sulfur as a potential power plant FGD byproduct. Marketing advantages over sulfuric acid include over three times the equivalent sulfur concentration; nonhazardous, noncorrosive properties; easy stockpiling characteristics; less market competition in many locations; and recoverable energy, in the production of sulfuric acid, of approximately 8 MBtu/short ton of sulfur.

The market for FGD byproduct sulfur is wider and potentially greater than the market for FGD byproduct sulfuric acid. Sulfur can be marketed as a raw material to all acid plants regardless of their size and production costs. FGD byproduct sulfuric acid marketing is typically limited to relatively small high-production-cost acid plants.

Table 2. SEVEN POWER PLANTS MARKETING ACID IN THE 1983 $0.70 ACFL SOLUTION

Ranked order	Plant No.	Location	MW	Tons of acid	Consumer	Location	Tons of acid
1	0720000900	NC	1440	67,000	Royster	Norfolk, VA	15,000
					Swift	Norfolk, VA	26,000
					Weaver	Norfolk, VA	26,000
2	2455000250	KY	2011	339,000	Allied	Nitro, WV	101,000
					American Cyanamid	Hamilton, OH	71,000
					Army Ammun. Plant	Radford, VA	125,000
					Marion	Indianapolis, IN	42,000
3	2185000900	TX	2820	156,000	Olin	Pasadena, TX	156,000
4	1145001900	MI	800	75,000	American Cyanamid	Joliet, IL	38,000
					American Cyanamid	Kalamazoo, MI	7,000
					Swift	Calumet City, IL	30,000
5	4820001800	MI	2431	335,000	Dupont	North Bend, OH	131,000
					Dupont	Cleveland, OH	150,000
					Allied	Cleveland, OH	54,000
6	1655000300	FL	1280	197,000	Kerr-McGee	Cottondale, FL	11,000
					Royster	Mulberry, FL	186,000
7	0785000500	IL	550	74,000	USI Chemical	Tuscola, IL	74,000
				1,243,000			1,243,000

Although no byproduct sulfur sales were shown in either the 1983 or the 1978 model solutions a sensitivity analysis showed that it could become competitive in certain locations with a relatively small (10%) reduction in total FGD byproduct sulfur costs. At one power plant, for example, a reduction of 8.5% ($1.18 M/yr) of the total production cost ($13.91 M/yr) of FGD sulfur would result in a reduction in incremental cost sufficient to enter the market in competition with sulfur delivered from Port Sulphur.

Of the 16 power plants that have relatively low estimated sulfur production costs compared to limestone scrubbing, the first plant becomes competitive with a reduction in total FGD sulfur production costs of only 3.1%. The other 15 plants become competitive with reductions ranging from 5.4 to 23.7%. The combined production of these plants would be 265,723 short tons/year, equivalent to over 800,000 tons of sulfuric acid. This represents slightly over 3.0% of the projected 1983 sulfuric acid demand from sulfur-burning plants.

Marketing sensitivity to FGD byproduct sulfur is based on sulfur delivered from Port Sulphur at $70/long ton plus delivery and handling expenses escalated to 1983. At the level of projected 1983 Port Sulphur delivered costs, Canadian sulfur recovered from sour gas may be more competitive in the Midwest and Great Plains where 6 of the 16 plants analyzed are located.

CONCLUSIONS

The need of an extended lead time was apparent. The increasing period between the decision to build and plant completion requires a corresponding increase in projection lead time. It is critical, therefore, that projections be extended as far into the future as is practical. This may require more recent data than have been available through conventional channels.

The marketing potential of boilers projected for completion between 1978 and 1983 dominated the solution. Projections must emphasize future boilers since they are expected to be the most likely candidates for FGD byproduct marketing.

Although boilers reported to be in compliance are excluded from the model on the basis that strategy selection is final, this may not be necessarily true for future boilers if the strategy selected is clean fuel. Proposed regulations, higher clean fuel premiums, and scrubbing technology improvements could alter the choice of clean fuel as a total compliance strategy for future boilers. Model solutions should include future boilers if compliance is projected as a result of clean fuel usage.

Application of more stringent environmental controls must be analyzed for effects on model solutions.

As FGD, transportation, and sulfuric acid avoidable production costs increase, correspondingly higher ACFL values will be required to fully analyze potential market interactions for extended projections.

Greater emphasis should be given to potential transportation advantages of water traffic. Backhaul potential should also be considered where applicable.

FGD byproduct sulfur is not yet competitive with FGD byproduct sulfuric acid. A relatively small reduction (10% or less at nine power plants) in total FGD byproduct sulfur costs, however, could result in emergence of competitive FGD sulfur production.

Projections of avoidable construction costs for future sulfuric acid plants should be developed. The current model only considers the costs which can be avoided by shutting down an existing or future sulfuric acid plant operation. For future considerations it also includes all costs which can be avoided by not building new plants.

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5. O'Brien, W. E., and W. L. Anders. First Annual Update - Potential Abatement Production and Marketing of Byproduct Sulfur and Sulfuric Acid in the U.S. Tennessee Valley Authority, Office of Agricultural and Chemical Development, Muscle Shoals, Alabama. Prepared for the U.S. Environmental Protection Agency (in press).

LIMESTONE FGD OPERATION
AT
MARTIN LAKE STEAM ELECTRIC STATION

BY

MARK RICHMAN
MANAGER, PRODUCT IMPROVEMENT
RESEARCH-COTTRELL

ABSTRACT

The startup of the Flue Gas Desulfurization (FGD) System at Texas Utilities' Martin Lake Steam Electric Station, Unit #1, in April of 1977 marked the first use of FGD technology on a full-scale lignite installation. The station also features fully closed loop FGD operation with no liquid waste streams generated, disposal of the truckable FGD sludge-fly ash solid waste product at the mine site, and the use of bypass gases for all scrubbed gas reheating.

Since the startup, the Double-Loop$_{(TM)}$ FGD Systems for Unit #1 and Unit #2 (June of 1978) have operated at well in excess of 95% absorber efficiencies and 90% reagent utilization while consuming less than 1.3% of the station's electrical power production. In addition the FGD Systems have been extremely reliable, causing only 5 days of unscheduled boiler downtime combined as of December 1, 1978.

This paper, prepared by Research-Cottrell - the supplier of the FGD and Solids Handling Systems, presents a discussion of the Martin Lake S.E.S. FGD Systems design, startups, operating experience to December 1, 1978, performance and operating requirements.

DESIGN CRITERIA

Martin Lake Steam Electric Station consists of four 750,000 kilowatt lignite-fired units. The station, located in Rusk County, Texas, is surrounded on three sides by Martin Lake, which serves as a source of makeup and cooling water. The station is also less than ten miles from the source of its lignite fuel.

The Texas lignite fuel for the Martin Lake S.E.S. has the following characteristics:

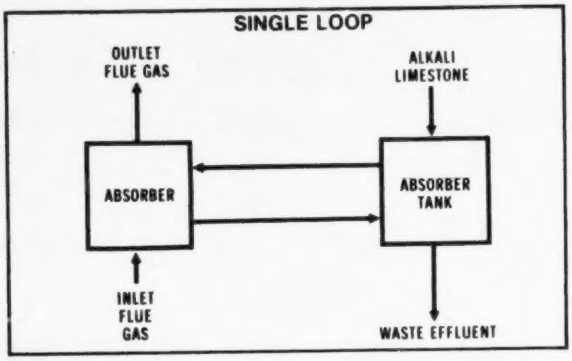
	MINIMUM	AVERAGE	MAXIMUM
Heating valve, BTU/LB	6,972	7,380	7,894
Ash Content, %	5.6	8.0	13.2
Sulfur Content, %	0.5	0.9	1.5
Moisture content, %	29.0	33.0	37.9

To produce 750,000 kilowatts of power with this fuel, each unit requires approximately 1,000,000 lbs/hr of lignite fired. To meet this demand, the utility operates a drag line with a 94 cubic yard bucket at the mine source.

The particulate emissions code for Martin Lake is 0.1 lbs/MMBTU. A 99.4% efficiency Research-Cottrell Double-Deck Electrostatic Precipitator is provided to achieve this requirement. The SO_2 emissions code for Martin Lake is 1.2 lbs/MMBTU. This computes to about 71% SO_2 removal for the worst case (i.e. maximum sulfur) fuel. Other requirements for the FGD System included: totally closed loop operation; a minimum of 25^{o}F reheat at the stack entry; and the production of a truckable, dumpable disposal product that can be transported back to the mine site.

FGD SYSTEM DESIGN

The process provided by Research-Cottrell for SO_2 removal at Martin Lake is the Double-Loop$_{(TM)}$ Limestone System. A schematic of the basic process design is shown in Figure #1. The Double-Loop$_{(TM)}$ process differs from simple single-loop processes in that two separate sets of chemical conditions are maintained. In a single loop process, shown below, the limestone slurry contacting the flue gas is kept at a fixed set of chemical conditions (i.e. solids composition and pH) for the required level of SO_2 removal. Since high levels


SINGLE LOOP
OUTLET FLUE GAS
ALKALI LIMESTONE
ABSORBER
ABSORBER TANK
INLET FLUE GAS
WASTE EFFLUENT

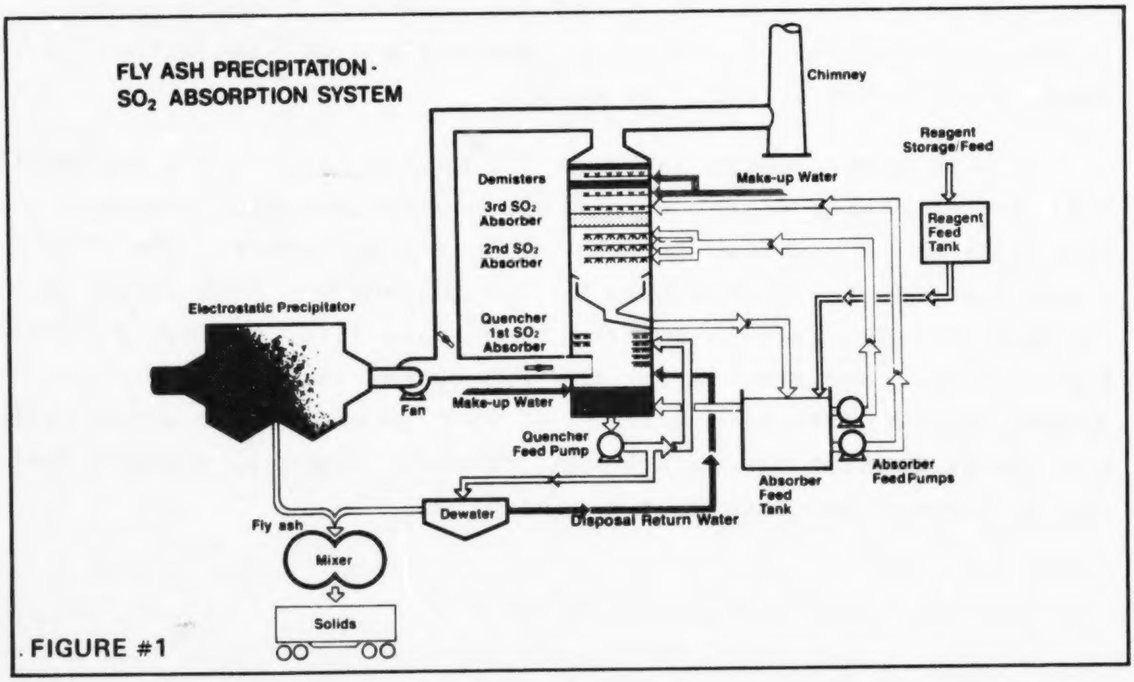
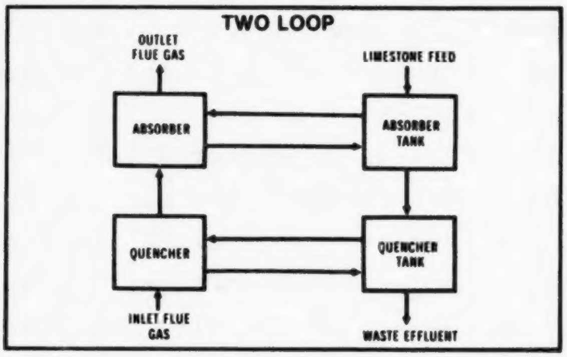

FLY ASH PRECIPITATION - SO2 ABSORPTION SYSTEM
Chimney
Reagent Storage/Feed
Reagent Feed Tank
Demisters
Make-up Water
3rd SO2 Absorber
2nd SO2 Absorber
Quencher 1st SO2 Absorber
Electrostatic Precipitator
Fan
Make-up Water
Quencher Feed Pump
Absorber Feed Tank
Absorber Feed Pumps
Dewater
Disposal Return Water
Fly ash
Mixer
Solids

FIGURE #1

of SO_2 removal efficiency can only be obtained with sufficiently high inventories of limestone reagent present, and since the solids blowdown to the solids handling/dewatering system is at the same chemical conditions as the scrubbing slurry, it is difficult to achieve both high SO_2 removal efficiencies and high reagent utilization with this type of system. As a result, typical first generation systems of this type achieve reagent utilization levels no better than 70 to 80%

In a two loop process, the first slurry loop (i.e. the Absorber Loop) operates identically as the single loop process operates to insure high SO_2 removal efficiencies. The second slurry loop (i.e. the Quencher Loop) receives the slurry discharge from the first loop and re-uses it to obtain reagent utilization levels exceeding 90%, even with absorber tower SO_2 removal efficiencies exceeding 95%.

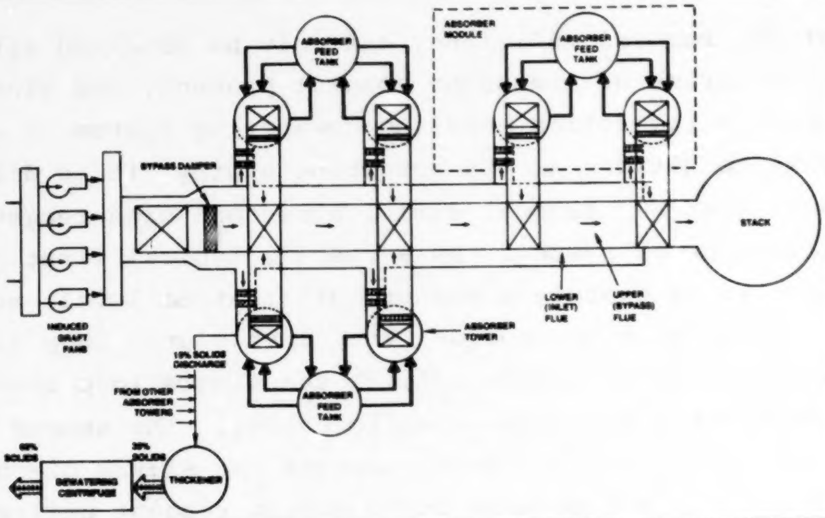


Using the Double-Loop$_{(TM)}$ approach, six absorber towers each were provided for Martin Lake #1 and for Martin Lake #2. A schematic of the arrangement of the absorber area is shown in Figure #2. Each absorber tower is designed to treat 12.5% of the maximum boiler flue gas output. Therefore the entire FGD System can treat up to 75% of the maximum boiler output with all six towers in operation. In order to achieve the 71% overall SO_2 removal requirement for the worst fuel case, this means that each tower must be capable of achieving 95% SO_2 removal efficiency.

The Martin Lake absorber tower design is presented in Figure #3. Each absorber tower, 28 feet in diameter by 100 feet tall, features three stages of SO_2 removal units followed by a two stage mist

FIGURE #2

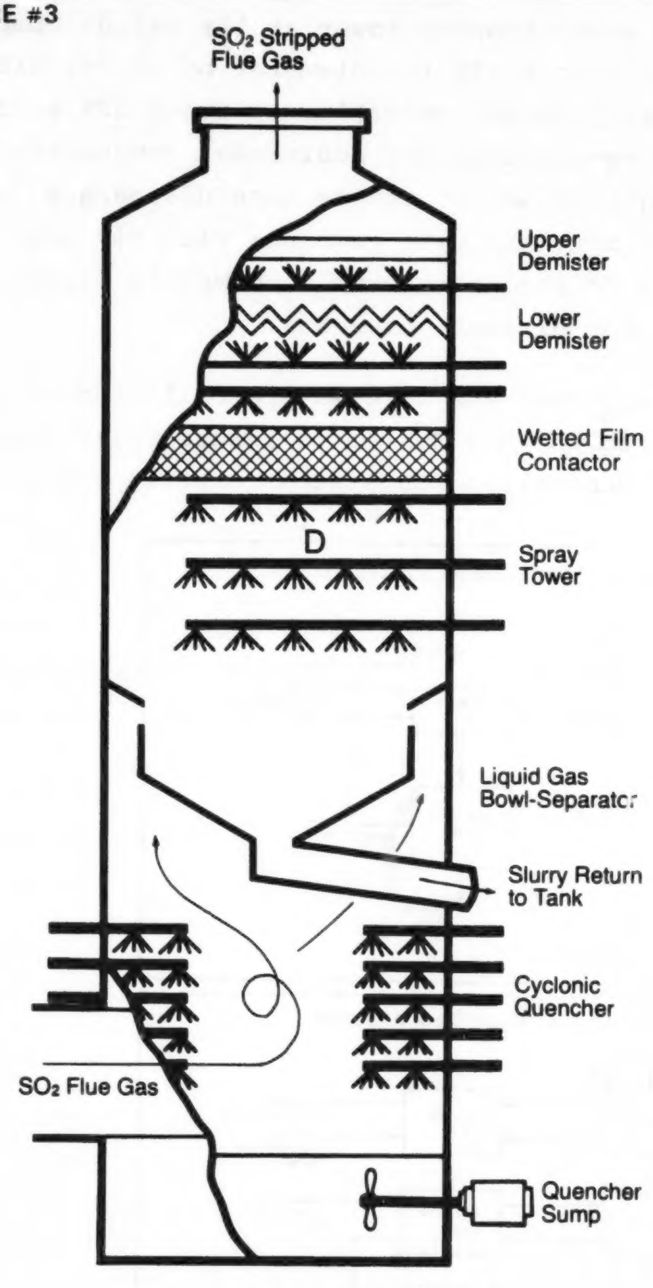
MARTIN LAKE #1 FGD SYSTEM ABSORBER AREA ARRANGEMENT


ABSORBER FEED TANK
ABSORBER MODULE
ABSORBER FEED TANK
BYPASS DAMPER
STACK
LOWER (INLET) FLUE
UPPER (BYPASS) FLUE
ABSORBER TOWER
INDUCED DRAFT FANS
15% SOLIDS DISCHARGE
FROM OTHER ABSORBER TOWERS
ABSORBER FEED TANK
60% SOLIDS
30% SOLIDS
DEWATERING CENTRIFUGE
THICKENER

elimination system. Boiler flue gas at 335°F enters the tower tangentially into the cyclonic quencher, a co-current spray chamber feeding quencher (i.e. second slurry loop) slurry. Here the gas is quenched to saturation, most of the fly ash not removed in the ESP is removed, and 25-30% of the SO_2 is removed. After passing through a liquid-gas separator which recycles absorber slurry back to the Absorber Feed Tank, the gas passes through three levels of absorber sprays and two feet of the fixed bed wetted film contactor. These two SO_2 removal sections remove virtually all of the remaining SO_2 by contacting the flue gas with absorber (i.e. first slurry loop) slurry. The cleaned gas then passes through two chevron mist eliminator stages and exits the absorber tower.

Once the cleaned gases exit the absorber towers, they are recombined with the portion of the boiler flue gases that was not scrubbed to reheat the treated gases. Since no more than 75% of the gases is treated at any one time, a minimum of 48°F of reheat (well above the required 25°F) is obtained. Note that at lower sulfur levels, where less than 71% overall SO_2 removal is required, less gas is treated so that more gas is bypassed giving even greater levels of reheat.

FIGURE #3

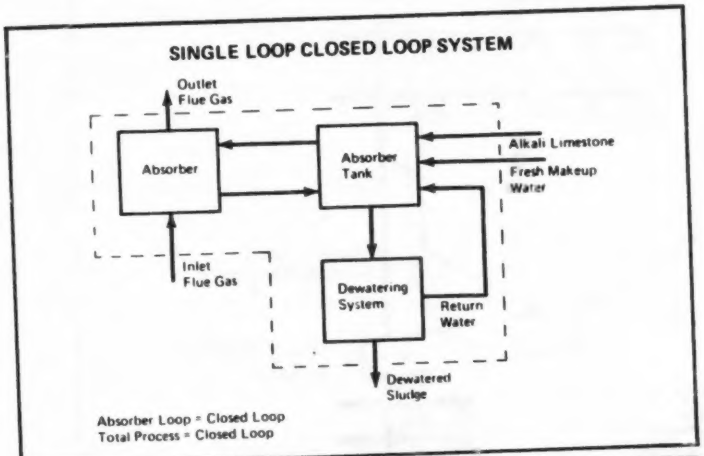

SO_2 Stripped Flue Gas
Upper Demister
Lower Demister
Wetted Film Contactor
D
Spray Tower
Liquid Gas Bowl-Separator
Slurry Return to Tank
Cyclonic Quencher
SO_2 Flue Gas
Quencher Sump

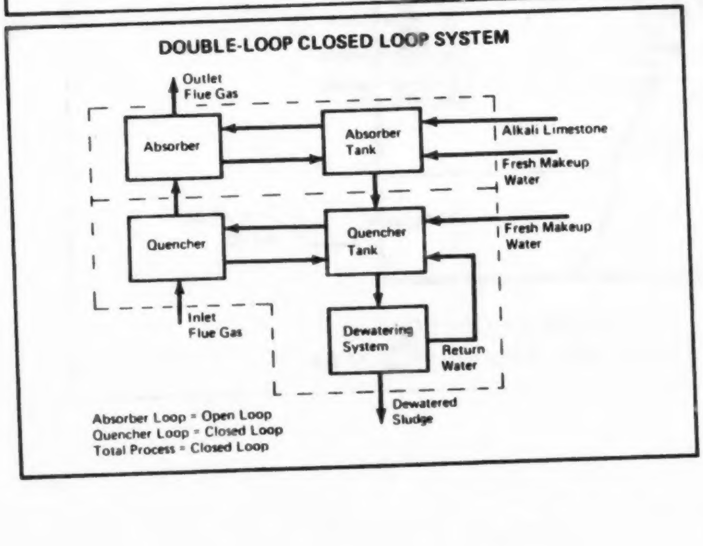
DOUBLE-LOOP(TM)
QUENCHER-ABSORBER TOWER

SOLIDS HANDLING SYSTEM DESIGN

Waste slurries from each absorber tower at 15% solids combine into a single stream to feed a 140 ft. diameter by 12 ft. sidewall gravity thickener. The thickener underflow at about 35% solids is fed to one of three centrifuges for additional dewatering. Cake from the centrifuges at 68-70% solids then discharges into a Muller-type blender. Here the cake combines with fly ash collected in the ESP to to form a truckable, dumpable blend. Railcars receive the blend for ultimate disposal.

Thickener overflow and centrate streams are collected in wet wells for recycle back to the FGD System to insure fully closed loop operation. First generation FGD systems experienced process


SINGLE LOOP CLOSED LOOP SYSTEM
Outlet Flue Gas
Absorber
Absorber Tank
Alkali Limestone
Fresh Makeup Water
Inlet Flue Gas
Dewatering System
Return Water
Dewatered Sludge
Absorber Loop = Closed Loop
Total Process = Closed Loop


DOUBLE-LOOP CLOSED LOOP SYSTEM
Outlet Flue Gas
Absorber
Absorber Tank
Alkali Limestone
Fresh Makeup Water
Quencher
Quencher Tank
Fresh Makeup Water
Inlet Flue Gas
Dewatering System
Return Water
Dewatered Sludge
Absorber Loop = Open Loop
Quencher Loop = Closed Loop
Total Process = Closed Loop

problems due to closed loop operation. As shown below, these single loop systems to not isolate the return water to any part of the process, but instead expose the entire process to disposal return water.

The Double-Loop$_{(TM)}$ closed loop system works differently in that disposal return streams are concentrated only in the Quencher Loop. The Absorber Loop, with the wetted film contactor and demisters, runs as an open loop, even though the total process is closed loop. As a result, the Absorber Loop can operate unsaturated in calcium sulfate, minimizing the potential for harmful scale formation.

STARTUP AND OPERATING HISTORY

Research-Cottrell provided the equipment, design engineering and advisory services during construction, checkout and startup for the FGD System. H. B. Zachry Co. constructed the FGD System. TUGCo was responsible for checkout of equipment and instrumentation, startup, operation and maintenance of the FGD System.

A chronological account of the first nineteen months of FGD System operating at Martin Lake is presented below. Following this we will discuss mechanical and instrument related operating problems and solutions.

April 21, 1977 - Flue gas admitted to tower 1C-T200.

April 26, 1977 - Flue gas admitted to tower 1C-T300.

April 29, 1977 - Flue gas admitted to tower 1B-T200 and 1B-T300; stack SO_2 monitor indicates the unit is in SO_2 compliance.

June 5, 1977 - Boiler shutdown for tube leak permits completion of construction on A module towers. Inspection of towers reveals wetted film contactors and demisters with little or no solids accumulation.

June 10, 1977 - Boiler back on line at 800-810 MW with all six towers now in operation.

June 24, 1977 - Preliminary acceptance tests at 750 MW indicate system can achieve compliance with four towers up to 900ppm SO_2 and with five towers up to 1,200ppm SO_2. Tower SO_2 removal efficiencies at 98-99%.

August 7, 1977 - Boiler operation resumed after month-long shutdown resulting from furnace damage.

August 16, 1977 - Slurry introduced to the thickener; thickener overflow returned to FGD System.

August 25, 1977 - Sludge introduced to centrifuges; blending operation begins.

September, 1977 - Environmental Protection Agency certifies Martin Lake #1 as being in full compliance with all air quality regulations.

November 3, 1977 - Due to low absorber tower gas flows and an inability to fully isolate the towers, the boiler was shut down. Inspection of the scrubbers revealed wetted film contactor (WFC) support beams in tower 1C-T200 had failed with WFC falling into the liquid-gas separator and tower sump. The inspection also showed tower inlet wet-dry line buildup sufficient to have contributed to the low gas flows.

November 8, 1977 - Boiler and all six towers restarted with all WFC removed to prevent additional support beam failures.

NOTE THAT THIS FIVE DAY PERIOD WAS THE ONLY BOILER SHUTDOWN PERIOD DURING THE FIRST NINETEEN MONTHS OF BOILER OPERATION CAUSED BY FGD SYSTEM MALFUNCTION.

(Note also that two other shutdowns (March 15, 1977 for three days and April 8, 1977 for two days), were related to bypass damper failure.)

January 28, 1978 - Tower 1A-T200 was the first to replace FRP support beams with alloy WFC support beams; other towers subsequently received alloy support beams. Inspection of this tower after 2½ months of operation with no WFC revealed significant demister buildup not present when the tower was operated with the contactor. This buildup necessitated the replacement of much of the demister packing and some demister beams. Note that an inspection of a tower with two feet of WFC amde in the Spring, 1978 after seven weeks of operation revealed the WFC and demisters to have only insignificant buildup.

April 2, 1978 - Unit #1 taken off line for annual two week outage.

May 1, 1978 - Unit #1 taken off line due to generator problems; boiler operation resumed on June 21, 1978.

June 22, 1978 - Unit #2 startup begins with 2 towers (B module) placed into service.

June 23, 1978 - Unit #2 C module placed into service.

August, 1978 - Unit #1 FGD System acceptance tests conducted by Texas Utilities; results show the Absorber Towers to be removing 98-99% of the incoming SO_2.

October 11, 1978 - Tower 1B-T200 taken out of service for 2 weeks to repair lining; Unit #2 remained in compliance during this period with the other 5 towers in operation.

October 19, 1978 - Unit #1 FGD System accepted by Texas Utilities

November 23, 1978- Unit #1 taken off line for month long scheduled outage.

FGD OPERATING PROBLEMS AND CORRECTIVE ACTION

The following discussion presents the major mechanical, structural and instrumentation related problems encountered during the first nineteen months of operation of the Martin Lake #1 FGD System and the first five months of operation of the Martin Lake #2 FGD System. Devices and pieces of equipment not discussed can be assumed to have performed satisfactorily, but should not be assumed to have been totally trouble-free.

Dampers - Dampers provided for the Martin Lake FGD System include a single-louver bypass damper, two consecutive louver dampers at each tower inlet, and single link-louver dampers at each tower out-let. Even with the addition of seal air flowers between the two tower inlet dampers, isolation of individual towers for personnel access and maintenance has been extremely difficult, but is now possible. Areas of malfunction have included bearings, seal strips, and linkages.

Expansion Joints - All gas-side expansion joints originally provided have been replaced due to the original joints' deterioration and inability to withstand exposure to liquids. The replacement joints are performing adequately to date.

Liquid Flow Switches - Sonic flow switches, originally provided to monitor flow deviations in key process loops, were found to be inadequate for electric generating station environments. They were eliminated, where possible, or replaced with magnetic flow meters.

pH Meters - Submersion devices in the quencher instrument wells experienced calibration drifts and electrical problems. These were replaced, following an extensive test program, by flow-through devices with ultrasonic cleaners. These have performed extremely well for 6 months.

Liquid Level Devices - Sonic level devices in the quencher instrument wells experienced problems related to electronics, poor calibration, slurry foaming, fouling of transducers and excessive liquid level variations. These problems were addressed and the devices have functioned adequately since October, 1977.

Gas Flow Meters - Poor gas distribution at the absorber tower inlets led to the relocation of these devices to the tower outlets. High instrument air moisture content and probe pluggage also contributed to poor operating performance.

On-Off Valve Operators - Electric operators on reagent feed and density control valves experienced numerous failures in early operation. Some of these were replaced with pneumatic valve operators and others were modified to improve performance.

WFC Support Beams - The original FRP support beams were replaced with 316SS beams in January and February, 1978. The FRP beams had failed completely in one tower and were on the verge of failure in several other towers when the WFC was removed from all towers on November 7 and 8, 1977.

Failure of these beams was due to inadequate structural design but was accelerated due to solids accumulations in the wetted film contactor. These solids accumulations were caused by operating the FGD System out of chemical balance due to problems with pH devices, on-off valve operators and the difficulty in manual control that these malfunctions created. Poor slurry distribution to the contactor may also have accelerated this buildup. As stated earlier, recent operation with two feet of WFC and with the new valve operators has indicated the WFC and demister packings to have had only insignificant buildup.

Wet-Dry Line - Buildup at the cyclonic inlets of the absorber towers was discovered in November, 1977. Model testing determined modifications to the inlets to allow operation of the towers for a full year without cleaning. These modifications have been successful and are being added to all absorber towers.

FGD SYSTEM PERFORMANCE

Several test programs were conducted by Research-Cottrell for TUGCo to evaluate the Martin Lake FGD System and to optimize operating setpoints. The following summarizes key results of the testing:

SO_2 Removal - Towers with four feet of wetted film contactor achieved greater than 99% SO_2 removal. Towers with no wetted film contactor achieved 80-85% removal at peak tower gas velocities. Based on this testing, two feet of WFC was installed in each tower and the maximum tower gas throughput was increased by 10%. Unit #1 acceptance tests in August of 1978 revealed absorber tower SO_2 removal to be greater than 98% with two feet of WFC at design velocities.

Reagent Utilization - Typically greater than 90%; average of 90-92% utilization. Tests on tower 1C-T200 on in situ forced oxidation showed that the tower can be operated with a reagent utilization in excess of 98% on a continuous basis.

Scale Control in Closed Loop Operation - Testing of the saturation levels of sulfate in each loop indicated an average Absorber Loop sulfate saturation level ∿ 0.90 and an average Quencher Loop sulfate saturation level ∿ 1.20. This indicates that, even during closed loop operation, the Absorber Loop operated unsaturated in SO_4, preventing sulfate scale formation in the absorber section of the towers. The Quencher Loop sulfate level shows, as expected, the sulfate level to be super-saturated but below the 1.3 times saturation critical level.

Disposal % Solids - Each system operated within 1 - 2% of design % solids with the thickener underflow ∿ 35% solids, the centrifuge cake ∿ 68% solids and the fly ash sludge blend at ∿ 82% solids.

Tower Pressure Drop - The absorber tower pressure drop at the design gas flow rate was within 0.1 IWC of the predicted value at 4.5 IWC. This does not include the two tower inlet louver dampers.

Degree of Oxidation - During normal operation, the quencher discharge has a sulfite:sulfate ratio of about 50:50 (or 1:1). A test program, using in situ forced oxidation on tower 1C-T200, produced 100 tons of solids with a sulfite:sulfate ratio of 1:99.

FGD SYSTEM OPERATING REQUIREMENTS

The following are operating and maintenance requirements for Martin Lake #1 FGD and Solids Handling Systems based on the first nineteen months of operation. Consumption rates are based on 750MW, 0.9% sulfur and 80% boiler availability per year. (Note that the unit normally operates at 800 - 820MW.)

Reagent Consumption - 240 tons/day or 70,000 tons/year of 95% $CaCO_3$ limestone.

Makeup Water Consumption - 550 gpm or 700 acre-ft/year.

Sludge Generation - 500 tons/day or 150,000 tons/year of 68% solids centrifuge cake.

Blended Waste Product - 13 railcar loads/day or 3,800 railcar loads/year.

Power Consumption - 10MW or 1.3% of station generated power - This includes ID Fan horsepower to account for FGD System ΔP, quencher pumps, spray tower pumps, WFC pumps, slurry agitators, reagent preparation system, solids handling system, scrubber control room operation, HVAC and lighting and all auxiliary equipment and motor control centers related to the FGD System. The Electrostatic Precipitator is not included in this total.

Operating Personnel - 2 Scrubber Area Operators/shift; 1 Reagent Area Operator/day (1 Shift); 2 Solids Handling System Operators for two units/shift.

Therefore, for Unit #1 FGD and Sludge Systems, a total of 3 1/3 Operators per shift are required.

Maintenance Personnel - Actual maintenance manpower has averaged 1,700 manhours/month. Actual chemical technician manpower has averaged 200 manhours/month. This computes to about 12 manmonths/month for maintenance and chemical technicians.